TITLE

A STUDY TO EVALUATE AND COMPARE THE PRODUCTION COSTS OF LIQUID FUELS FROM COAL AND FROM OIL SANDS IN ALBERTA

FINAL REPORT

BY

DYNAWEST PROJECTS LTD.

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TABLE OF CONTENTS

Page

1. SUMMARY 1

2. INTRODUCTION 5

2.1 General 5
2.2 Design Basis for Oil Sands Plant 6
2.3 Design Basis for Coal Liquefaction Plant 6

3. THE PRODUCTION OF LIQUID FUELS FROM OIL SANDS 18

3.1 Introduction 18
3.2 The Mining Operation 18
3.3 Bitumen Extraction 22
3.4 Primary Upgrading 25
3.5 Secondary Upgrading 27
3.6 Utilities 31
3.7 Offsite Facilities 35
3.8 Environmental Impact 37

4. THE PRODUCTION OF LIQUID FUELS FROM COAL USING THE H-COAL PROCESS 45

4.1 Introduction 45
4.2 The Mining Operation 45
4.3 Coal Storage and Preparation 48
4.4 Coal Hydrogenation, Product Separation and Upgrading 49
4.5 Ancillary Process Operations 53
4.6 Utilities 57
4.7 Offsite Facilities 59
4.8 Environmental Impact 61

5. PROCESS ECONOMICS 66

5.1 Key Economic Summary 66
5.2 Description of Process Costs 66

6. PROCESS COMPARISONS 77

6.1 Introduction 77
6.2 Basic Ore Treatment 77
6.3 Product Hydrotreating and Product Quality 80
6.4 Hydrogen Production 83
6.5 Gas Treatment 84
6.6 Sulphur Recovery 84
6.7 Effluent Treatment 84
6.8 Utilities 84
6.9 The Environmental Impact 86
6.10 Product Utilization 88

TABLE OF CONTENTS (Cont'd)

Page

7. REFERENCES 109

8. APPENDIX 114

LIST OF FIGURES

1.1 Pictorial Representation of Capital Costs

3.1.1 Liquid Fuels from Oil Sands. Overall Block Diagram
3.1.2 The Production of Liquid Fuels from Oil Sands. Overall Process Schematic.
3.2.1 Single Bench Dragline Operation. Plan.
3.2.2 Single Bench Dragline Operation. Range Diagram.
3.2.3 Single Bench Dragline Operation with Rehandle.
3.2.4 Double Bench Dragline Operation.
3.3.1 Flow Schematic of Extraction and Froth Treatment Plants.
3.3.2 Bitumen Recovery Versus Fines in Extraction Feed Mineral.
3.3.3 Extraction and Froth Treatment Material Balance.
3.4.1 Fluid Coker Flow Schematic.
3.5.1 Material and Energy Balance Flowsheet. Upgrading Operations.
3.5.2 Naphtha Hydrotreater.
3.5.3 Distillate/Gas Oil Hydrotreater.
3.5.4 Gas Recovery Unit.
3.5.5 Amine Treater for LPG.
3.5.6 Merox Unit for LPG.
3.5.7 Fuel Gas Treater.
3.5.8 Three Stage Sulphur Plant (Claus Unit).
3.5.9 SCOT Tail Gas Cleanup.
3.5.10 Hydrogen Plant (Coke Gasification).
3.5.11 Block Flow Diagram. Oxygen Plant.
3.5.12 Power Plant Schematic.

4.1.1 Liquid Fuels from Coal. Overall Block Diagram.
4.1.2 The Production of Liquid Fuels from Alberta Coal. Overall Process Schematic.
4.2.1 Single Dragline Operation.
4.2.2 Tandem Operation. Position of Second Machine.
4.3.1 Coal Storage and Preparation.
4.4.1 Coal Liquefaction. Unit 200. Hydrogenation and Fractionation.
4.4.2 H-Coal Reactor
4.4.2.1 Naphtha Hydrotreater.
4.5.1.1 Light Ends Processing.
4.5.1.2 Conventional Amine Unit.
4.5.2.1 Reforming and Shift Train.
4.5.2.2 Gasification Train.
4.5.2.3 Selexol Plant - Hydrogen Unit.
4.5.3 Oxygen Plant.
4.5.4.1 Emission Control System
4.5.4.2 Claus Sulphur Plant.
4.5.4.3 Sulphur Plant Tail Gas Treating. Beavon Process.
4.5.5.1 Effluent Control System.

LIST OF FIGURES (Cont'd)

5.1 Pictorial Representation of Capital Costs

6.1 Boiling Point Ranges of Oil Sands and H-Coal Synthetic Crude

LIST OF TABLES

1.1 Capital and Operating Costs
1.2 Major Products and Byproducts

2.2.1 Synthetic Crude from Oil Sands
2.2.2 Properties of Oil Sands Synthetic Crude Naphtha
2.2.3 Properties of Oil Sands Synthetic Crude Kerosene
2.2.4 Properties of Oil Sands Synthetic Crude Diesel
2.2.5 Properties of Oil Sands Synthetic Crude Heavy Fuel Oil
2.3.1 Picardville Coal Deposits
2.3.2 Comparison of Wyoming and Canadian Coals
2.3.3 Petrographic Analysis of Picardville Coal
2.3.4 Synthetic Crude from H-Coal Process

3.1 Oil Sands Component Material Balance
3.6.1 Utitities Summary for an Oil Sands Plant

4.1 Coal Liquefaction Component Material Balance
4.2.1 Major Equipment Schedule for 19 Million t/a Coal Mine (Surface Strip Mining)
4.6 H-Coal Process. Major Utilities Summary

5.1 Summary of Major Products
5.2 Summary of Capital Investment
5.3 Summary of Operating Costs
5.4 Escalation Figures
5.5 Oil Sands Process. Direct Field Costs
5.6 H-Coal Process. Direct Field Costs
5.7 Coal Process Capital Investment. Mine Process Split
5.8 Operating Costs. Mine Process Split

6.2.1 Alternatives to the Clark Hot Water Extraction Process
6.2.2 Processes for the Primary Upgrading of Bitumen
6.3.1 Synthetic Crude from Oil Sands
6.3.2 Raw Synthetic Crude from H-Coal Process
6.3.3 Properties of Oil Sands Synthetic Crude Naphtha
6.3.4 Properties of Oil Sands Synthetic Crude Kerosene
6.3.5 Properties of Oil Sands Synthetic Crude Diesel
6.3.6 Properties of Oil Sands Synthetic Crude Heavy Fuel Oil
6.3.7 Experimental Data for Hydrotreating H-Coal Derived Naphtha
6.3.8 Properties of H-Coal Derived Syncrude used in Hydrotreating Studies
6.3.9 Properties of H-Coal Derived Gas Oils used in Hydrotreating Studies
6.3.10 Properties of Hydrotreated Naphthas from Oil Sands and Coal Liquefaction Plants

LIST OF TABLES (Cont'd)

6.3.11 Properties of Hydrotreated Distillate/Turbine Fuel from Oil Sands and Coal Liquefaction Plants (vs. Jet A Spec).
6.3.12 Properties of Hydrotreated Distillate/Turbine Fuel from Oil Sands and Coal Liquefaction Plants (vs. Diesel Specifications)
6.3.13 Experimental Data for Hydrotreating H-Coal Derived Gas Oil
6.3.14 Properties of Hydrotreated Gas Oil from Oil Sands and Coal Liquefaction Plants
6.9.1 Land Area Required for Synthetic Fuel Projects
6.10.1 Probable Disposition of H-Coal Products

1.0 SUMMARY

Dynawest Projects Ltd. has been commissioned by CANMET to evaluate and compare the production costs of liquid fuels from coal and oil sands in Alberta. An earlier interim study$^{(1)}$ concluded that the coal based liquefaction route that should be used in the comparison is the Hydrocarbon Research Inc. (HRI), H-Coal process.

This report contains descriptions of both the oil sands and coal based processes for the production of liquid fuels. Material and energy balances are presented for each process together with process schematics. Basic economic data in the form of capital cost estimates and annual operating costs are presented for plants producing 4 830 000 t/a of liquid synthetic crude oil products, these are summarized in Table 1.1. Within the constraints of this study and excluding infrastructure costs, we have estimated that the capital investment required for the coal based route exceeds that for the oil sands process by approximately $0.75 billion. The key reasons are illustrated in Figure 1.1 which shows relative as well as absolute values. The major difference can be seen to lie in the primary ore treatment process which in the H-Coal process comprises multi-train coal preparation and liquefaction equipment costing about twice as much as its counterpart, the extraction and primary upgrading units, in the oil sands process.

The more sophisticated coal technology requires a larger operating and maintenance staff. It also has a higher chemical demand (principally natural gas for power production and catalyst for the liquefaction reaction). These combine to give a higher annual operating cost of about $270 million.

The major products and byproducts produced by each process are summarized in Table 1.2.

The contract under which this study was performed stipulated that the full economic analysis using the data presented in this report will be performed by CANMET.

From Table 1.1 it can be seen that the production of liquid fuels is more costly from coal than from oil sands. This straight forward cost penalty is offset by a number of factors:

1) The coal process produces a higher proportion of the more valuable lighter petroleum fractions.
2) The coal process has less of an environmental impact.

3) Coal deposits are not located in such remote regions as oil sands deposits so that the social impact of any coal based project would be far less than that of an equivalent oil sands project.

It should also be noted that the higher capital cost of the coal process is concentrated in the liquefaction reaction. It is this part of the coal process that is subject to the most intensive research and in which costs may be expected to fall as development proceeds.

Table 1.1 **Capital and Annual Operating Costs ($ x 10^6)**

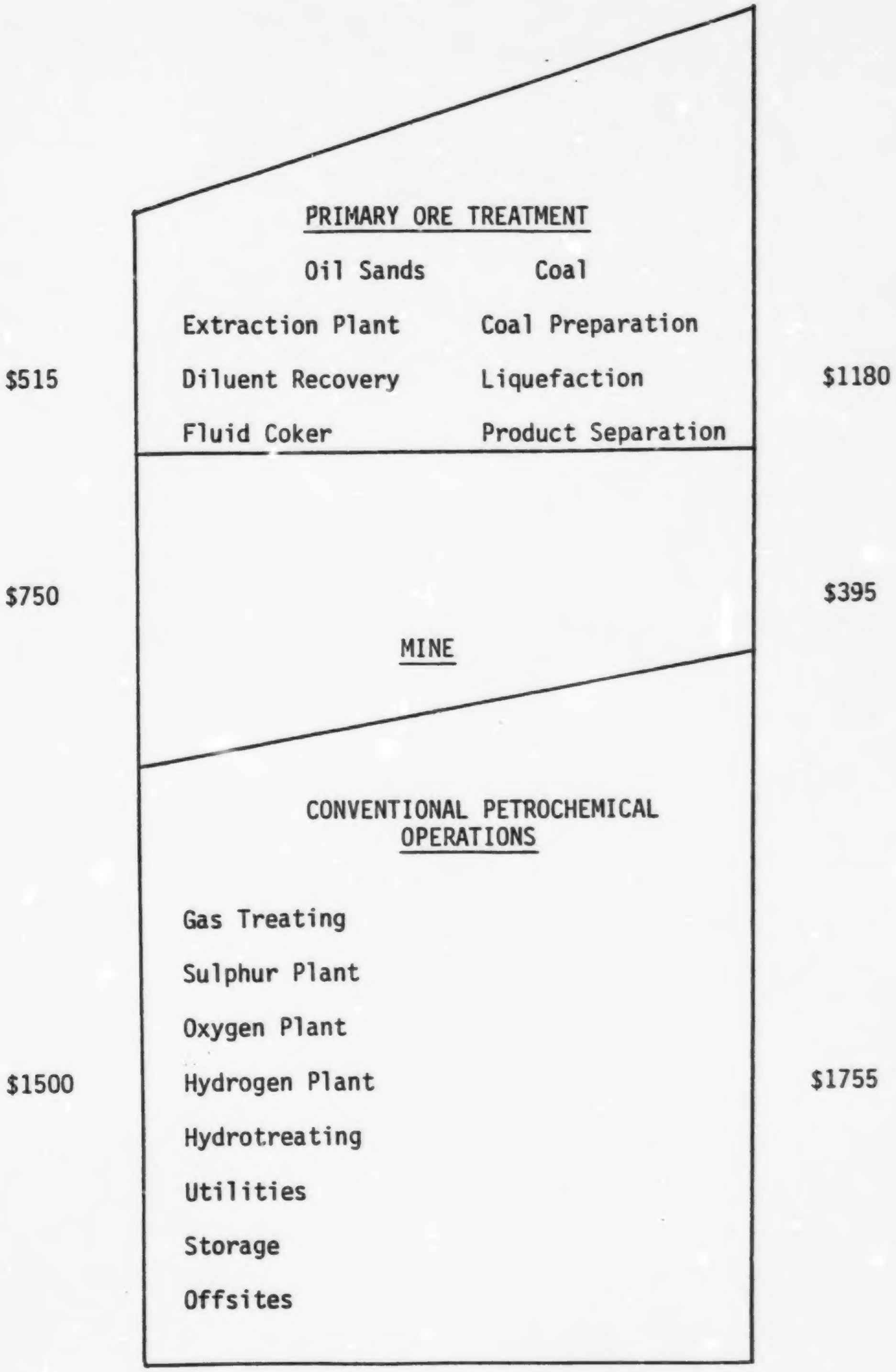
	Oil Sands	Coal Liquefaction
Capital		
Direct Field Costs	2 767	3 331
Indirect Field Costs	1 878	1 986
Preproduction Costs	918	887
Post Start-Up Costs	180	-
Working Capital	102	198
Interest During Construction	1 500	1 669
Subtotal	7 345	8 071
Major Infrastructure Costs	963	-
TOTAL	8 308	8 071
Operating (per annum)		
Wages and Salaries	257	324.4
Operating Materials, Chemicals	56.6	161.6
Maintenance Materials	89.7	152.9
Corporate Overhead	24.8	33.5
Property Tax and Insurance	40.0	56.2
Reclamation and Revegetation	6.0	5.5
Imported Power	3.9	15.9
TOTAL	478.0	750.0

Table 1.2 Major Products and Byproducts (kt/a)

	Oil Sands	Coal Liquefaction
Naphtha	1 029	1 910
Distillate/Turbine Fuel	1 126	1 810
Gas Oil	2 675	1 110
LPG	110	0
Sulphur	280	50
Ammonia	0	173
Phenols	0	19

FIGURE 1.1

PICTORIAL REPRESENTATION OF CAPITAL COSTS


PRIMARY ORE TREATMENT
Oil Sands
Coal
Extraction Plant
Coal Preparation
$515
Diluent Recovery
Liquefaction
$1180
Fluid Coker
Product Separation
$750
$395
MINE
CONVENTIONAL PETROCHEMICAL OPERATIONS
Gas Treating
Sulphur Plant
Oxygen Plant
$1500
Hydrogen Plant
$1755
Hydrotreating
Utilities
Storage
Offsites

OIL SANDS	COAL
Direct Costs	Direct Costs
$2765 million	$3330 million
Total Costs	Total Costs
$7345 million	$8070 million

2.0 INTRODUCTION

2.1 GENERAL

This report was commissioned to provide a comparison on an equal technical basis of the production of liquid fuels by the liquefaction of a defined coal feedstock with the production of a product of similar specification by a current generation surface mined oil sands operation. The synthetic crude from each process has to be suitable as a refinery feedstock for the production of transportation and heating fuels.

Where possible information has been taken from sources available to the public. Nonconfidential data obtained from HRI, the licensor of the H-Coal process is included in the report.

The H-Coal process was selected for reasons detailed in (1) but which were based on it being either equal to or better than its competitors in the following key areas:

1. Equivalence of synthetic crude product to that from an oil sands plant.
2. Control of product slate.
3. Proven capacity on sub-bituminous coal.
4. Potential process economics.
5. Available published data.

The scale of operations in each case has been selected as 16,000 t/d (operating day) (approximately 120,000 bbl/d) of liquid synthetic crude product so as to be comparable with the existing Syncrude operation. This scale of operations also matches reasonably well with the estimated coal reserves for the Picardville coal defined by CANMET as the feedstock for this study. In defining both plants as having the same daily production rate, there is an implicit assumption that both have the same overall availability and reliability to give the same annual production rate. Since so many operations are common to both processes (e.g. surface mining, gasification, hydrotreating, sulphur recovery, effluent treatment, power production) it would be difficult to justify an alternative decision in a preliminary study of this nature.

Key reference documents described in the design basis sections below are only referred to subsequently when specific figures or tables are taken from them.

One of the criteria for this study was that, where possible, information should be obtained from publicly available documents.

2.2 DESIGN BASIS FOR OIL SANDS PLANT

The basic reference documents for the whole of the oil sands process have been the applications and supplemental information documents issued by the Alsands consortium(Ref. 2 - 12). This choice was dictated by the fact that the application documents form a consistent set of information unlike those for the Syncrude and Suncor plants which underwent many changes during the course of the projects.

Some changes have been made to the case presented in the Alsands application.

1. The scale of operations has been reduced from 18,250 t/d to 16,000 t/d. (To make the size consistent with the largest installed operating plant i.e., Syncrude).
2. The original Alsands application showed the use of gas-oil, one of the main liquid products, as a plant fuel. This was rejected by the ERCB(11). Our flowsheet, therefore, shows all plant fuel provided by the gaseous and LPG byproducts.
3. The Alsands application shows initial hydrogen production from natural gas with a change after 5 - 8 years to production from fluid coke. Our flowsheet simply describes the ultimate operation using fluid coke as the hydrogen source.

It should also be noted that the overall process efficiency is based on the Alsands assumption of an average extraction plant ore feed content of 0.121 mass fraction bitumen and 0.145 mass fraction of fines. The key operation determining overall efficiency is the extraction process and the effect of different ore concentrations is discussed in Section 3.3 (Bitumen Extraction).

The overall crude composition is given in Table 2.2.7. Properties of the main fractions are given in Tables 2.2.2, 2.2.3, 2.2.4 and 2.2.5.

2.3 DESIGN BASIS FOR COAL LIQUEFACTION PLANT

The description of the surface strip mining operation is based on the Highvale mine located about 100 km SW of Picardville, Alberta. The Highvale mine will have an ultimate capacity of 14 million t/a comparable to the 19 million t/a required for the mine supplying the liquefaction plant. The main source of data for the mine description is the application of Calgary Power Ltd. to the ERCB, references (13) and (14) and a further reference (15) describing the development of the mine.

In basing the description on the Highvale mine it was assumed that major design factors (e.g. overburden depth, spoil heap stability, etc.) will be the same at Picardville. In the absence of a detailed survey of Picardville, this assumption is unavoidable. However, since both coals are sub-bituminous Plains coals, both are strip mineable and the locations are relatively close, this does not seem too unreasonable an assumption.

The description of the H-Coal liquefaction process is, at the recommendation of HRI, the process licensor, based on a study completed by Fluor for EPRI (The Electric Power Research Institute). This study(16) examined two coal feedstocks, one of which, Wyodak sub-bituminous coal is similar, but not identical, to Picardville coal. The main properties of Picardville coal as supplied by CANMET are given in Table 2.3.1, a comparison with Wyodak coal is given in Table 2.3.2. A petrographic analysis of the coal, Table 2.3.3, was also provided to HRI who were able to make some preliminary comment. These were:

1. The mean reflectance of the Picardville coal (0.40) is higher than that of Wyodak (0.36) indicating a slightly higher rank coal.
2. The H/C ratio is lower.

These observations led HRI to the conclusion that reaction conditions would probably have to be more severe to obtain the same yields as for Wyodak coal.

Because no experimental data is available for Picardville coal it has been assumed that yields are the same as for Wyodak coal. Clearly they will not be exactly the same but specific differences can only be established by experimental work.

The one area in which we have modified the material and energy balance from(16) is by accommodating the different production rates of the two principal byproducts sulphur and ammonia to reflect the different sulphur and nitrogen levels in the two coals. Reference (16) describes the production of synthetic crude with much higher residual sulphur and nitrogen levels than the upgraded product obtained from an oil sands plant. To bring the coal derived synthetic crude fractions up to the same quality hydrotreaters have been added on each major liquid product stream. The data for the hydrotreating of the H-Coal synthetic crude fractions was obtained from references (17) and (18).

The composition of the synthetic crude product is given in Table 2.3.4 which identifies the breakdown of the three main synthetic crude components. These may be exported separately or blended depending on the refining strategy adopted which, because of the site's proximity to Edmonton, could well be different to the strategy adopted for the oil sands product.

While there is little doubt that upgraded coal liquids of comparable quality to oil sands derived synthetic crudes can be produced the range of H-Coal liquids for which experimental hydrotreating data is available is, as yet, limited. We have therefore deliberately overestimated the hydrogen consumption required to produce products of equivalent value to the oil sands fractions. As can be seen from the detailed mass balance (Section 4) total hydrogen consumption is relatively insensitive to the hydrotreating demand, which amounts to only about 15% of the total.

Table 2.2.1 Synthetic Crude from Oil Sands

Crude Quality

API	33.0
Sulphur (mass fraction)	2×10^{-3}
Nitrogen (mass fraction)	0.7×10^{-3}

Composition (Volume Fraction)

C_3	0.01
C_4	0.038
C_5 - 150°C	0.175
150 - 335°C	0.375
335°C+	0.402
	1.000

Table 2.2.2 (27) **Properties of Oil Sands Synthetic Crude Naphtha**

Boiling range	C_5 - 150°C
Density, kg/m^3	697
Sulphur, mass fraction	1×10^{-6}
GC Hydrocarbon analysis, volume fraction	
Paraffins	0.69
Naphthenes	0.25
Aromatics	0.06
Benzene	0.011
Toluene	0.027
Xylene	0.02
Research Octane No.	58
Motor Octane No.	57

Table 2.2.3 (27) Properties of Oil Sands Synthetic Crude Kerosene

	Synthetic	Jet A Specifications
Density, kg/m^3	830	775-839
Viscosity, cSt at 40°C	1.3	-
Freeze pt., °C	-60	-40 max
Sulphur, mass fraction	38×10^{-6}	2×10^{-3} max
Nitrogen, mass fraction	4×10^{-6}	-
Hydrogen, mass fraction (D 3701)*	0.129	-
Aromatics, volume fraction (D 1319)*	0.32	0.20 max
Naphthalenes, mass fraction (D 1840)*	0.006	0.03 max
Smoke point (D 1322)*	13	20 min
Luminometer no. (D 1740)*	29.3	45 min

* ASTM Test methods

Table 2.2.4 (27) **Properties of Oil Sands Synthetic Crude Diesel**

Density, kg/m^3	897.4
Viscosity, cSt at 40°C	3.3
Cloud point, °C	-27
Pour point, °C	-50
Sulphur, mass fraction	300×10^{-6}
Nitrogen, mass fraction	25×10^{-6}
Aromatics, volume fraction (D 1319)	0.44
Cetane no. (measured) (D 613)	34
Cetane index (calculated) (D976)	42
Distillation (D 86)	
IBP°C	157
10%	192
50%	274
90%	329
FBP	334

Table 2.2.5 (27) Properties of Oil Sands Synthetic Crude Heavy Fuel Oil

Boiling range, $^{\circ}$C	455°C+
Yield on crude, volume fraction	0.13
Density, kg/m^3	936
Viscosity, cSt at 100°C	18
Sulphur, mass fraction	2×10^{-3}
Nitrogen, mass fraction	1.2×10^{-3}
Pour Point, $^{\circ}$C	9
Conradson carbon residue, mass fraction	2×10^{-3}
BMCI	49

Table 2.3.1 Picardville Coal Deposits

Rank (ASTM):	Sub-bituminous C	
Location:	40 Km North of Edmonton	
Reserves:	Strip mineable - 399 million tonnes Underground - 18 million tonnes	
Mineability:	typical plains conditions	
Environmental:	water supply adequate good agricultural land	
Infrastructure:	Rail, highways, towns and cities, clearly zoned.	
Typical Assay:		
Proximate		
	Capacity Moisture -	29.3%
	Ash (as received)-	6.8%
	Fixed Carbon -	41.2%
	Calorific Value -	7 830 BTU/lb
Ultimate (MAF)		
	Carbon -	75.5%
	Hydrogen -	4.7%
	Nitrogen -	1.4%
	Oxygen -	17.9%
	Sulphur -	0.5%

Table 2.3.2 Comparison Wyoming and Canadian Coals

	Wyoming Sub-bituminous Coal Wyodak Mine			Canadian Sub-bituminous C Coal		
	AR	MF	MAF	AR	MF	MAF
Moisture, wt%	31.34	-	-	29.3	-	-
Proximate Analysis, wt% (dry basis)						
Ash	7.08	7.08	-	6.8	6.8	-
Volatile Matter	46.17	46.17	49.69	52.0	52.0	55.79
Fixed Carbon	46.75	46.75	50.31	41.2	41.2	44.21
Ultimate Analysis, wt% (dry basis)						
Carbon	66.58	66.58	71.65	70.37	70.37	75.5
Hydrogen	4.93	4.93	5.30	4.38	4.38	4.7
Nitrogen	1.05	1.05	1.13	1.30	1.30	1.4
Sulphur	1.14	1.14	1.23	0.47	0.47	0.5
Oxygen	19.22	19.22	20.68	16.68	16.68	17.9
Ash	7.08	1.08	-	6.80	6.80	-
HHV, Btu/lb	7,900	11,506	12,382	7,830	11,075	12,253
H/C (Molar Ratio)	0.89			0.75		

Notation

AR -	as received
MF -	moisture free
MAF -	moisture and ash free

Table 2.3.3 Petrographic Analysis of Picardville Coal

Mean Reflectance	0.40	
	Composition	
Huminite	Ulminite A	10
	Ulminite B	40
	Densimite	21
	Gelinite	7
	Corpohuminte	7
Total Huminite		85
Inertite	Semi-fusinite and fusinite	7
	Other Inertite	4
Total Inertite		11
Liptinite		4
Total		100

Note: Huminite is the vitrinite equivalent in low rank coals.

Table 2.3.4 Synthetic Crude from H-Coal Process

	Naphtha	Turbine Fuel	Distillate Boiler Fuel
API	58	32	20
Sulphur (mass fraction)	4×10^{-6}	20×10^{-6}	1×10^{-6}
Nitrogen (mass fraction)	1×10^{-6}	20×10^{-6}	20×10^{-6}
Mass fraction of Total Synthetic Crude	0.4	0.37	0.23
Composition (Mass Fraction)			
C_4	0.01	-	-
C_5 - 40°C	0.07	-	-
40 - 95°C	0.23	-	-
95 - 150°C	0.45	-	-
150 - 205°C	0.24	-	-
175 - 315°C	-	1.00	-
260 - 425°C	-	-	1.00
	1.00	1.00	1.00
FIA Volume Fraction			
A	0.19	-	-
O	0.00	-	-
P & N	0.81	-	-
RON	67	-	-
BMCI	-	-	65

3.0 THE PRODUCTION OF LIQUID FUELS FROM OIL SANDS

3.1 INTRODUCTION

The process for the production of liquid fuels from oil sands is simply summarized on Figure 3.1.1. This shows a mine producing 200,000 t/d of oil sand which are fed to an extraction plant to remove 22,000 t/d bitumen. This bitumen is processed in primary and secondary upgrading units to produce 16,000 t/d of synthetic crude. Fuel gas, LPG and sulphur are produced as byproducts. The major support units are a hydrogen plant which produces hydrogen from the primary upgrading coke byproduct, a sulphur plant and a utilities plant supplying all site demands.

The operation is shown in considerably more detail on Figure 3.1.2. The flow rates and compositions for the streams shown on this figure are presented in Table 3.1. Detailed descriptions of each operation in the process are given in the sections that follow. Note that in the more detailed figures presented in subsequent sections all flowrates are referred back to Table 3.1.

3.2 THE MINING OPERATION

Two alternative basic techniques have been employed for the surface mining of Albertan oil sands. These are the use of bucketwheels as operated by Suncor and the use of draglines as operated by Syncrude and proposed by Alsands. This study assumes the use of draglines, it enables a more direct comparison with the coal mining case.

In essence, the mining operation consists of the following stages:

1. Site clearing;
2. Surface drainage;
3. Muskeg removal;
4. Subsurface dewatering;
5. Dragline mining of overburden and oil sand;
6. Ore and waste handling and transportation;
7. Tailings management and site reclamation.

Each stage is described below.

3.2.1 Site Clearing

The areas which comprise a typical oilsands lease are covered with a mixture of saleable trees and bush and other non-saleable growth. The useful trees are removed and the remaining vegetation stripped and burned.

3.2.2 Surface Drainage

Surface water from undisturbed areas adjacent to dewatered areas is collected by interceptor ditches and diverted around the site. The muskeg areas have to be drained prior to removal. This is achieved by the provision of shallow drainage ditches to the surrounding area.

3.2.3 Muskeg Removal

After drainage, muskeg is removed during the winter by trucks and front end loaders. (Winter removal is dictated by the inability of front end loaders to handle and transport muskeg unless it is frozen). The muskeg is stored in special compounds for future reclamation purposes and to prevent it from flowing over the countryside on thawing.

3.2.4 Subsurface Dewatering

A complex system of subsurface aquifers exists in many oil sands leases. The Alsands application, for example, identifies four such aquifers. Each has to be depressurized (if necessary) and drained prior to mining. The aquifers can vary in their potential impurities (principally bitumen or chloride) and therefore in the methods available for their disposal. In the Alsands case, the following methods are all used:

1) Direct to natural drainage;
2) Via the tailings disposal pond;
3) To natural drainage via settling pond;
4) By subsurface injection.

3.2.5 Dragline Mining of Overburden and Oil Sand

Four draglines are used to excavate both overburden and oil sand. Single bench mining is carried out to a total depth of 70 m and two bench mining to a depth of 90 m. When mining oil sands at these depths, there is no technical requirement for pre-stripping of overburden. However, front end loaders and trucks can be used for prestripping as required to supplement dragline productivity.

During mining operations, the ability of the dragline to mine selectively is exploited so as to reduce the amount of lean ore delivered to the extraction plant.

Factors to be considered when defining the precise operating method to be employed by the draglines include:

1) Overburden depth;
2) Stable angle of highwall slope (which decreases with highwall depth);
3) Stability of spoil heaps;

Since the overburden depth varies throughout the mine, a number of different dragline configurations are required. Figures 3.2.1 and 3.2.2 show typical single bench operation. All of the centre reject and some of the overburden are cast directly back into the mine. However, with the overburden depth shown in Figure 3.2.2, some of the overburden has to be rehandled because of the limiting spoil heap angle. This material is stored on the original surface and has to be rehandled before the oil sand can be stored there. This requirement leads naturally to the operation of draglines in pairs on the same level as shown in Figure 3.2.3. The first machine is chopcutting, producing a temporary waste rehandle stockpile and initiating the cast waste pile. The second machine is mining conventionally, completing the cast waste pile and producing the oil sand stockpile.

Figure 3.2.4 shows how the operation is modified for two bench operation. The upper bench is limited to 35 m. The upper dragline produces a waste rehandle stockpile and an oil sand stockpile. The lower machine only produces an oil sand stockpile, all waste material is cast directly back into the mine.

3.2.6 Ore and Waste Handling and Transportation

All installed and proposed oil sands mines use conveyor systems for the bulk of material transportation. A project of this magnitude will typically have two main 2.4 m wide trunk conveyors, each capable of handling all the extraction plant feed.

Oil sand is initially transferred to the trunk conveyor using bucketwheel reclaimers to feed 1.8 m wide face conveyors, which in turn feed the trunk conveyors. There is one bucketwheel reclaimer/face conveyor combination per dragline (ie. four in total).

In principle off-line oil sand conveyors can be used for handling waste material. However, the practical and logistical problems imposed by this approach generally necessitate the use of an independent system, usually based on the use of trucks.

3.2.7 Tailings Management

The tailings produced by the Hot Water Extraction Process consist of a clay, silt, sand, bitumen and water slurry. For the early years of operation, this effluent is stored within tailings dykes and then subsequently in the mined out area.

Two methods of tailings disposal are used, dyke building and overboarding. The dyke building method ejects the tailings at the crest of the existing dykes. As the tailings flow to the centre of the tailings pond, the coarse sand rapidly settles and the remaining slurry flows to the middle of the pond. The sand that deposits on the dyke is mechanically compacted and used to build up the dyke. The overboarding technique is used during the winter months. The slurry is discharged directly into the liquid within the pond. The sand particles settle on a widespread area forming a broad beach while the water and fine solid particles flow into the centre of the tailings pond. No mechanical compaction is used.

There are two reasons for using the overboarding method during the winter:

1) The tailings are emitted at approximately 70°C and the steam produced in the cold winter air prevents the safe operation of bulldozers that compact the solids.
2) A broad base is required to support the new dykes that are constructed during the summer.

The coarse sand that is contained in the tailings settles very quickly. The remaining sludge slowly separates, over a two to three year period, into two phases, a clear water layer and a gel like sludge containing 0.2 to 0.3 mass fraction solids. The gel like layer is extremely fragile and settles with time of the order of decades. As there is a large amount of water used in the Hot Water Extraction Process, the volume of tailings that must be stored is very high.

The volume of sludge that must be contained varies with the production rate of the facility but tailings ponds of up to 30 km^2 with a depth of 40 m (19) would be typical for the proposed facilities.

3.3 BITUMEN EXTRACTION

The Clark Hot Water Extraction Process is used to separate bitumen from the Athabasca Oil Sand into a form suitable for upgrading. The process schematic is shown in Figure 3.3.1 and is comprised of the following major steps:

1. Plant Stockpile and Feed System;
2. Conditioning;
3. Primary Separation;
4. Scavenging;
5. Froth Treatment.

3.3.1 Plant Stockpile and Feed System

The stockpile and reclaim system is located in front of the Extraction Plant. This serves as a surge to delink the mine and Extraction Plant operations.

Oil sand is delivered from the mine by conveyors and either stockpiled by stackers at the entrance of the Extraction Plant or delivered directly to surge bins as plant feed. The stockpiled oil sand is reclaimed by a bucketwheel reclaimer to suplement the Extraction Plant feed during periods when feed requirements exceed mine output. The oil sand in the surge bins is discharged onto apron feeders, weighed and fed by four inclined conveyors to four parallel processing lines for extraction. The oil sand feed rate to the extraction facilities is 197,200 t/sd of 0.121 mass fraction bitumen. At this grade, the average fines content will be approximately 0.145 mass fraction.

3.3.2 Conditioning

In the conditioning step, oil sand is combined with hot water at 85oC and agitated in tumblers (rotary drums) to ablate the oil sand lumps. Caustic is added to control the alkalinity of the pulp at a pH of 8 to 8.5 and steam is injected to maintain the temperature. The conditioned pulp is screened at the discharge end of the tumblers to separate the oversize material, consisting essentially of clay lumps and rocks, from the dispersed bitumen, sand, clay and water. The oversized reject is transported to the mined-out area or the tailings pond for disposal.

3.3.3 Primary Separation

Separation of the dispersed bitumen from the conditioned pulp is accomplished in a separation cell by diluting the pulp with hot water and recycled middlings (the central layer in the separation cell), then holding it in a quiescent state. The use of middlings recycle reduces the hot water make-up requirement, which is governed by the amount of fines in the oil sand feed. The water rate is adjusted to control the middlings density/viscosity. The coarse sand particles settle rapidly to the conical bottom of the separation cell while the bitumen rises forming a frothy layer on top of the cell. This layer, referred to as primary forth, is skimmed off and pumped to the froth treatment plant. The sand in the cell bottom is continuously raked to a central discharge cone from which it is pumped after dilution to the tailings pond.

The unsettled fines, consisting essentially of silt and clays from the oil sand, tend to accumulate in the middlings. The rate of rise of bitumen flecks is retarded due to increase in viscosity in the cell caused by this accumulation. The fines must be continually withdrawn from the central layer in the cell in the form of a dilute middlings slurry; otherwise accumulation of fines will cause the viscosity to buildup until separation of the bitumen ceases.

The middlings withdrawn contain significant quantities of bitumen which can be economically recovered in the scavenging circuit.

3.3.4 Scavenging

Conventional floatation cells are employed in the scavenging operation. Air is injected into the middlings which is further mechanically agitated to promote the formation of a bituminous froth. The recovered froth is sent to a froth settler to improve its quality. It is then combined with froth from the primary separation and sent to froth treatment.

Tailings from the scavenging operation are combined with the primary separation tailings for disposal to the tailings pond.

3.3.5 Froth Treatment

In the Froth Treatment Plant, the combined froth from the primary separation and scavenging operations is treated to provide a suitable feedstock for the Upgrading Plant. The froth is first heated then deaerated prior to dehydration and demineralization by dilution centrifuging.

Dilution centrifuging utilizes crude naphtha, a recycle material, as a diluent to reduce the specific gravity and viscosity of the bitumen, with the result that the entrained solids and water, which are heavier than diluted bitumen, can be segregated.

The diluted bitumen is centrifuged in the primary solid-liquid separators (scroll centrifuges) at relatively low gravitational forces to remove the large mineral particles. The high energy secondary solid-liquid-liquid separators (disc centrifuges) further reduce the mineral and water content.

Diluted bitumen is then stored in surge tanks prior to feeding it to the Upgrading facility.

The Froth Treatment Plant tailings will be pumped separately to a localized area of the tailings pond. This will maximize the residence time of this stream so that more fines will settle prior to water recycle. It also facilitiates the potential future installation of equipment to recover either bitumen or heavy minerals both of which are more concentrated in this stream than in primary tailings.

3.3.6 Recovery

The recovery efficiency is directly related to the fines content of the plant feed as indicated in Figure 3.3.2. The estimated average recovery efficiency for the Extraction Process is 91.6 percent, excluding diluent losses, for an average fines content in mill feed of 0.145 mass fraction. However, naphtha losses in the centrifuge plant will reduce the total hydrocarbon recovery to 90.3 percent.

3.3.7 Material Balance

The material balance shown in Figure 3.3.3 gives the average stream day flow rates to produce 21,723 m^3/sd of bitumen. It is based on an average mill feed grade of 12.1 percent bitumen and an average fines content of 14.5 percent.

3.4 PRIMARY UPGRADING

The Primary Upgrading Facilities incorporates the following process units in two parallel trains:

1. Diluent Recovery Units;
2. Fluid Coking with CO Boilers.

3.4.1 Diluent Recovery Units (DRU)

The purpose of the Diluent Recovery Units is to produce a dry bitumen from the diluent/bitumen/water mixture that is fed to them from the Extraction Plant and to recycle the recovered diluent back to the Extraction Units. The bitumen enters the Diluent Recovery Units diluted with an equal volume of naphtha and carries residual clay and water. The naphtha is produced in the Naphtha Hydrotreater and its make-up rate to offset Extraction Plant losses is about 410 m^3/sd.

The feed rate to each Diluent Recovery Unit is approximately 20,750 m^3/sd and the separation of water, naphtha and bitumen in a process is achieved in a carefully staged heating and flashing operation. Recovered water is collected and routed to the sour water treating facilities and the dry bitumen product flows via intermediate storage to the Fluid Cokers.

3.4.2 Fluid Coking

Fluid Coking is employed to upgrade the bitumen. This process was first commercialized in 1954 and conceptually the process is similar to the universally-employed Fluid Catalytic Cracking Process except that catalyst is not used. A fluidized bed of coke granules formed in the coking process is used in place of the catalyst. The unit configuration is shown on Figure 3.4.1.

Bitumen feedstock from the Diluent Recovery Units is fed to the reactor. The subsequent coking or cracking reaction produces a wide range of hydrocarbon products and causes additional coke to form in the fluidized bed

The coke bed is fluidized by the addition of steam at the reactor bottom and by the product vapours. The vapour products pass overhead through cyclone separators which remove most of the entrained coke for return to the reaction zone. The vapours are further scrubbed and cooled with oil, which is recycled to the reactor and then fractionated into the desired product cuts for further processing. Coke, which has given up considerable heat to feed preheat and cracking is circulated back to the coker burner where partial combustion occurs with air. Hot coke is returned to the reactor from the burner to heat balance the process.

To maintain constant fluid bed inventory, product coke is removed. After quench cooling and some internal particle classification, the coke is transported to storage.

The Fluid Coker is limited to a maximum product end point dictated by the maximum operating temperature on the overhead scrubber. Higher temperatures would result in unacceptable coke build-up. Coker gas and naphtha are routed to the Gas Recovery Unit, while light and heavy oil streams are sent to Hydrotreaters. There is a minimum of intermediate storage due to the unstable nature of the Coker product streams.

The CO Boilers act as the incinerator for both Fluid Coker burner gas and the tail gas from the Sulphur Plant and generate about 160 t/h of steam on each unit. The flue gas from the burner contains carbon monoxide and sulphur dioxide. The CO Boilers also provide Diluent Recovery Unit heat. Fines entrained through the burner cyclones to the CO Boilers are removed by electrostatic precipitators on the CO Boiler flue gas. The fines from the electrostatic precipitators are collected and transferred to the coke storage hoppers.

The sulphur level of the coke burned in the burner is lower than in either the liquid feed or in the net coke. Preferential burning of carbon over sulphur at the burner conditions results in an accumulation of sulphur on the coke rejected to storage. The net product coke from the burner is cooled in the quench elutriator by a stream of water and steam. The vapour produced tends to sweep the small coke particles back into the burner, resulting in a larger particle size for the net coke product. This is pneumatically conveyed to storage hoppers from where it can be transferred to storage, returned to the unit, or sent to the Coke Gasification Plant.

3.5 SECONDARY UPGRADING

The Secondary Upgrading Facilities incorporate the following process units:

1. Naphtha and Distillate/Gas Oil Hydrotreaters;
2. Gas Recovery Units and Treating Facilities;
3. Coke Gasification Units;
4. Claus Sulphur Plants and Tail Gas Recovery Units.

With the exception of the four gasification reactors, the Secondary Upgrading Facilities comprise two parallel trains.

The units for primary and secondary grading are arranged as shown on the Material and Energy Balance flowsheet on Figure 3.5.1.

3.5.1 Hydrotreaters

Hydrotreatment of various crude oil components is required to reduce sulphur, nitrogen, olefin and aromatic concentrations to acceptable levels for downstream refining operations.

The Naphtha Hydrotreating Process reacts hydrogen and naphtha in the vapour phase over a fixed bed catalyst. Butanes/butylene, coker naphtha and recycle naphtha from the Gas Oil Distillate Hydrotreaters make up the feed to two parallel Naphtha Hydrotreaters, each producing 2,530 m^3/sd of hydrotreated butane and heavier liquid. A Diluent Preparation Unit, common to both Naphtha Hydrotreaters, separates a narrow boiling range naphtha which is recycled to the Extraction Plant as diluent make-up. The remaining hydrotreated butane plus product is blended to synthetic crude. The naphtha is suitable for feed to conventional refinery Catalytic Reforming Units. Refer to Figure 3.5.2 for a typical naphtha hydrotreater.

The Distillate/Gas Oil Hydrotreating process combines hydrogen with distillates or heavy oils predominantly in the liquid phase over fixed bed catalysts, normally at higher pressures and longer residence times than for lighter feed.

The total volume of distillate and gas oil hydrotreated product is 14,400 m^3/sd. A characteristic of hydrotreating heavy oil fractions is the generation of a naphtha which is high in aromatics and low in other qualities. This is sent to the Gas Recovery Unit and subsequently to the Naphtha Hydrotreaters for treatment. This ensures that the final naphtha product meets product quality sulphur and nitrogen criteria. A small amount of propane and lighter gas produced is sent to the Gas Recovery Unit.

Recycle gas is scrubbed with amine solution to remove hydrogen sulphide and the rich amine is returned to a common regeneration system for sulphur recovery.

A typical Distillate/Gas Oil Hydrotreater is shown on Figure 3.5.3.

3.5.2 Gas Recovery/Treaters

The Gas Recovery/Treaters section collects gas and naphtha from the Fluid Coker, gas and naphtha from the Distillate/Gas Oil Hydrotreaters, gas from the Naphtha Hydrotreaters. The hydrocarbon feed is fractionated to produce the following products:

1. Ethane and lighter gas (fuel gas);
2. Liquid propane/propylene (LPG);
3. Liquid butanes, butylenes and naphtha.

A typical fractionator schematic is shown on Figure 3.5.4. The Gas Recovery fractionation heat requirements are met in part by a circulating stream of Fluid Coke fractionator gas oil. The liquids butanes, butylenes and naphtha are routed to the Naphtha Hydrotreater.

The LPG is transferred to an amine treating unit where sour gas is removed and fed to the Sulphur Plant. Then the LPG is caustic treated before sending it to the extractive Merox unit. The LPG is then routed to storage. Figures 3.5.5 and 3.5.6 show a typical amine treating and Merox unit.

The fuel gas stream is sent to the Fuel Gas Treater before sending it to the utilities and upgrading area as plant fuel. A typical gas treating unit is shown on Figure 3.5.7. At Syncrude the gas from the gas recovery section is sweetened with diethanolamine (DEA) to remove H_2S.

3.5.3 Sulphur Recovery

The acid gases from the Gas Recovery/Treaters and the Hydrogen Plant are sent to the Sulphur Plant. The Sulphur system consists of two parallel trains comprising three-stage Claus units followed by SCOT (Shell Claus Offgas Treating) units. These units normally operate on 50 percent of the plant throughput. The Claus units are designed for 96 percent sulphur recovery and when combined with the SCOT units, the overall Sulphur Recovery should be 99.9 percent. Each sulphur plant is designed for 60 percent of total production.

The conventional Claus process is widely used. Figure 3.5.8 shows a typical three stage ammonia burning Claus sulphur plant. The sour water stripper off gas is fed to the Claus Sulphur Plant and contains ammonia. In the Claus Sulphur Plant, the ammonia is destroyed(c). To convert H_2S to sulphur, SO_2 must be present. SO_2 is generated by burning about one third of the H_2S to provide the required stoichiometric ratio of H_2S to SO_2 for maximum sulphur production. Some of the main acid gas feed is fed to a separate combustion chamber with the ammonia rich sour water off gas. The combustion air is supplied by using air blowers. Because of the importance of sulphur removal, spare blowers are required. The combustion of H_2S must be controlled to provide the proper proportion of SO_2 to H_2S. This is achieved by limiting the amount of oxygen fed to the burner. Since ammonia and H_2S react to form ammonium sulphide, which dissociates at a temperature of approximately 70°C, it is important to keep this gas stream above that temperature to maintain H_2S and ammonia in the gaseous state. Cooler temperatures may result in salting out of the ammonium sulfide, which plugs up condenser tubes and pipe lines, requiring periodic flushing or cleaning out.

The SCOT process was developed in 1972 to recover sulphur compounds which have passed through the Claus Plant. These compounds are reduced over a cobalt/molybdenum catalyst to hydrogen sulphide which is then recovered in an amine absorber. The hydrogen sulphide stripped from the the amine is then recycled to Claus Plant feed. The tail gas is passed to the CO boiler for incineration of any residual hydrogen sulphide to sulphur dioxide. The hydrogen sulphide content of the tail gas ranges from 200 to 300 x 10^{-6} volume fraction. A typical SCOT process flowsheet is shown on Figure 3.5.9.

3.5.4 Sour Water Handling

Sour water containing ammonia and hydrogen sulphide is produced from the Diluent Recovery, Fluid Coker, Naphtha Hydrotreater, Gas Oil Hydrotreater and Gas Recovery Units. The sour water from the upgrading units is fed to the main sour water stripper. The Coke Gasification units have their own sour water strippers. The stripped gas is sent to the Sulphur Plant and the stripped water is routed to the recycle water storage basin at the rate of 6,385

m^3/d from the main stripper and 345 m^3/d from the Hydrogen Plant stripper (Coke Gasification Units).

3.5.5 Hydrogen Plant

To meet the demand of 4.17 x 10^6 Sm^3/d of hydrogen for the Hydrotreating units, the fluid coke is gasified in a Texaco gasifier. As high purity hydrogen is required, oxygen must be used instead of air. There is an oxygen plant that separates oxygen from air before passing the oxygen to the hydrogen plant.

The coke fed to the gasifier is slurried in water. Additives are required to stabilize the slurry which is then pumped to gasification pressure. Steam and oxygen from the air separation plant are also fed to the gasifier. Operating temperature in the gasifier is from 1300 to 1550°C at a pressure of up to 8370 kPa.

Reactions in the Gasifier produce a synthesis gas, rich in carbon monoxide and hydrogen. No hydrocarbons above methane are formed. The gas on leaving the gasifier is quenched with water and then deflected, thereby solidifying and separating most of the liquid slag from the gas. This ash, molten at the bottom of the reactive zone, flows into a quench pot at the bottom of the gasifier. In practice, an incomplete conversion of carbon is obtained. Typically 0.01 mass fraction of the carbon remains in the ash. At the quench pot the slag granulates and is removed via a water flooded lock hopper.

The gaseous product is cooled by raising steam and quenched in a water tower. H_2S is removed prior to the shift convertors in which carbon monoxide and steam are converted to carbon dioxide and hydrogen. The carbon dioxide is removed to produce the hydrogen feed to the hydrotreaters.

Soot and ash from the water quench leaves as a slurry in water. Sour gas from the stripper is sent to the sulphur plant. A typical Texaco Coke Gasification Plant is shown on Figure 3.5.10.

3.5.6 Oxygen Plant

Oxygen is required for production of high purity hydrogen in the gasification plant. Conventional cold box technology is used for the separation of oxygen from air. A schematic diagram is shown in Figure 3.5.11.

3.6 UTILITIES - GENERAL

The Utility Plant is designed to satisfy the total energy requirements of the Project. Since the power producing facilities are connected by a tie line to the provincial grid network, the power producing facilities will be internally controlled in order to achieve an acceptable import-export power balance while minimizing demands on the grid. Production and consumption of major utilities is summarized in Table 3.6.1.

3.6.1 Raw Water Storage and Filtration

Under normal operation with water recycle from the tailings pond, fresh water requirement is about 70,000 m^3/sd. Water is pumped from the Athabasca River and the river water requirements are minimized by using local surface run-off and subsurface water and by maximizing tailings pond recycle water. The water intake is located on the river bank. There is a settling reservoir adjacent to the Athabasca River equipped with a high head pump to lift the water to the plant site.

The water settling reservoir is a dam constructed in an abandoned channel in the Athabasca River. Suspended solids in the river water are removed in the settling reservoir. The water intake system at the bank has both low and high level ports for minimizing sediment intake to the system.

Athabasca River water, after settling, is processed by the water treatment facilities in two major sections. In the first section, the clarifier /filter processes the water and the processed water is divided into two streams. One stream goes to the Upgrading and Extraction plants while the other is used as demineralizer feed.

In the second section the filtered water passes through several stages of treatment including ion exchange in the demineralizer plant. The resulting treated water is suitable for use as makeup for the high pressure boiler feed water system. Deaerators are included as part of the boiler feed water treatment facilities. Boiler feed water is required for all steam generators in both Upgrading and Utilities.

Some water from the settling reservoir is used as recycle water make-up which is used for hot water production. Blowdown from the water treatment plant is neutralized to a pH range of 6 to 9 before being discharged to the tailings pond.

3.6.2 Boiler Feedwater

Boiler feedwater is generally used for production of high pressure steam. High pressure steam is mainly produced by the CO boiler. Other sources of high pressure steam are the hydrogen plant and utilities heat recovery boiler. Boiler feedwater is also required by the hydrogen plant for the gasifier feed fluid coke slurry. Some of the boiler feedwater is required to desuperheat the high pressure steam for use in the extraction plant conditioning drums when the steam provided by the upgrading units does not match demand.

3.6.3 Hot Water

Hot water is generated by passing the recycled water from the recycle water storage through upgrading coolers and condensers in the utilities plant. It is stored in the hot water surge tank before being directed to the conditioning drums in the extraction plant. Some of the hot water is recycled.

3.6.4 Recycled Water

The tailings pond is the main source of supply for the recycle water. The tailings pond collects water from extraction plant tailings, total blowdown from plant, subsurface aquifer water and rain water. Reclaimed water is directed to the recycle water storage basin. Process water blowdown and oil-free water from the oil separator are also fed to the recycle water storage basin. Some of the recycle water is used as cold water dilution for the main tailings from the extraction plant. Most of the recycled water is used for producing hot water.

3.6.5 Cooling Tower and Cooling Water System

The main supply of cooling water is from the recycled water storage. The recycled water removes heat from the upgrading area by using the upgrading coolers as a preheat for the recycled water. The recycled water is then heated further by the condensing steam heat exchangers before it is directed to the hot water surge tank for use in bitumen extraction.

Process and utility cooling towers remove heat from both the Process and the Utility sections. The utility cooling water tower is primarily for use with the condensing turbine. Cooling water makeup required for the utility cooling towers is 12,000 m^3/sd. Blowdown from the cooling towers is sent to the tailings pond.

3.6.6 Potable and Gland Water

Potable and gland water is supplied to the Extraction and Upgrading sections from the water treatment plant. The potable water is treated to Alberta Environment standards for municipal water supplies. Appropriate emergency storage is also included. The plant has a sewage treatment facility which may be combined with that of a local town.

3.6.7 Steam and Power Generation

Superheated steam is generated in both the Upgrading area and the Utility Plant. The CO boiler associated with the Fluid Coker produces the majority of the high pressure steam for the power plant. Most of the steam is used to generate power for the Mine, Extraction, Upgrading and Offsites. Low pressure steam is generated by the Utility plant for Extraction and Upgrading.

Electrical power is generated within the complex. Emergency startup and peak demands for power associated with the dragline operation, are purchased from the provincial grid. The integrated operation of gas turbines and back pressure turbo generators is designed to obtain a high operating efficiency from the steam and power systems. A simplified flow sheet of the power plant is shown on Figure 3.5.12.

The major categories of equipment involved in the Utility Plant are:

1. Gas Turbine-Driven Electric Generators;
2. Heat Recovery Steam Boilers;
3. Back-Pressure Steam Turbine-Driven Electric Generators;
4. Condensing Steam Turbine-Driven Electric Generators;
5. Ancillary Equipment.

3.6.7.1 Gas turbine-driven electric generators. The gas turbine system consists of an air compressor, a combustor, a hot gas expander and an electric generator. The compressor supplies compressed air to the combustor where a fuel air mixture is burned. The fuel used is fuel gas from the amine plant. For normal operation the fuel gas supplied by the Upgrading Section is more than sufficient. However, short term variations in fuel supply/demand are satisfied by adjusting the propane content of the fuel system. High pressure hot gas is exhausted through the expander turbine, driving both the electric generator and the air compressor. The high temperature exhaust gas from the turbine outlet, which is rich in oxygen, provides preheated combustion air to the downstream heat recovery boiler. There are three gas turbine electric generators of approximately 65 MW output capacity each. One of these is a spare unit during normal operation.

3.6.7.2 Heat recovery steam boilers. The heat recovery steam boiler recovers heat from the gas turbine exhaust and supplemental fuel is burned to control both superheat in the produced steam and the rate of steam production. The exit temperature of the flue gas from the heat recovery boiler is approximately 200°C.

There are three 5,200 t/d boilers of which two are used during normal operation with one spare. All boilers will be required when one of the two Upgrading trains is shutdown with the mine and Extraction Plant operating at design conditions.

The high pressure steam produced from the heat recovery boiler is used for generating electrical power in the back pressure and condensing steam turbine generators. Low pressure steam for extraction can be supplied by high pressure steam letdown when one or both of the upgrading units are off-line.

3.6.7.3 Back-Pressure steam turbine. The back-pressure steam turbines provide both electric power generation and low pressure steam for process use. This by-product power is generated at high efficiency as long as there is use for process steam. Electrical power requirements in excess of low pressure process steam requirements are more efficiently generated through condensing turbines.

There are three back-pressure units installed and one spare during normal operation. With one of the Upgrading trains shutdown and the mine and Extraction Plant at normal operation, all units will be required.

3.6.7.4 Condensing steam turbines. The condensing steam turbines are designed for vacuum conditions at the turbine steam exhaust. The efficiency of the condensing turbine is higher than that of the back-pressure unit due to the greater pressure drop through it. The thermal efficiency, however, is lower than that of the by-product system where the steam exhaust is used for process purpose. The lower thermal efficiency is due to the large quantity of heat in the exhaust steam being lost to the cooling water system. In the condensing service, only one 18 MW generator is driven.

3.6.7.5 **Ancillary facilities.** Process and Utility cooling towers remove heat from both the Process and Utility sections. The primary use of the utility cooling water tower is for cooling with the condensing turbine.

Plant gas compressors supply fuel gas to the gas turbines at the required pressure. Only gas turbine fuel is handled in these compressors.

Deaerator facilities are a part of the water treatment facilities for boiler feedwater preparation.

Steam trim exchangers provide steam heating of Extraction Plant water when heat from the process units is insufficent.

The boiler feedwater pumps provide boiler feedwater required for all steam generators in both Upgrading and Utilities.

Desuperheater stations provide flexibility in operation such as balancing of electrical production with process steam demands. The units produce steam at a suitable pressure and temperature, independent of the electrical production. This is achieved by blending higher pressure/temperature steam with treated water to produce the desired steam level. The units are not required during normal operation but only when the LP steam produced by upgrading operations does not meet the Extraction Plant demand.

3.7 OFFSITE FACILITIES

3.7.1 Tankage

Total tankage used for storage of diluted bitumen, dry bitumen, sour water, synthetic crude and miscellaneous items is about 950,000 m^3. Individual tankage capacities may differ from those shown below as a result of the individual manufacturer's operating preference. (For example, Suncor have concentrated on diluted bitumen storage, whereas Syncrude have concentrated on dry bitumen storage).

A typical storage capacity for diluted bitumen is about 10 days supply requiring tankage of 415,000 m^3. Two days of hold up is used for dry bitumen intermediate storage and the tankage capacity is 43,500 m^3. One weeks storage capacity is provided for sour water and two weeks for synthetic crude product. Tankage for sour water storage is 44,700 m^3 and for synthetic crude is 266,700 m^3. The remaining storage capacity of 185,000 m^3 is reserved for miscellaneous items such as chemicals, water, propane, upgraded crude components and off specification material.

3.7.2 Sulphur Storage and Handling

Liquid sulphur from the Sulphur Recovery Units is pumped from a local storage pit to the sulphur storage area. The sulphur is then formed into solid blocks.

3.7.3 Coke Storage and Handling

Solid coke from the Fluid Coker is slurried together with fly ash and coke fines from the electrostatic precipitator and pumped to the coke storage area before feeding it to the hydrogen plant. The electrostatic precipitator is located on the outlet of the CO boilers. Excess water is decanted and returned for re-use in the slurry process. Make up water is required to compensate for water entrained in the coke storage area. The storage area is lined to prevent seepage to ground water.

3.7.4 Flare and Blowdowns

All units in the facility are provided with appropriately sized pressure relieving devices and blowdown headers which are connected to the flare system. The flare system is provided with an automatic ignition system and knockout drums.

3.7.5 Potable Water and Sanitary Sewage

An in-plant system provides potable water for the camp and sewage facilities. The water is treated to Alberta Environment standards for municipal water supplies. Emergency storage is also installed. Sanitary sewage is treated in a Rotary Biological Contactor secondary sewage treatment plant at 95 percent efficiency to reduce the inlet BOD and suspended solids before discharge.

3.8 ENVIRONMENTAL IMPACT

3.8.1 General

Some disruption of the environment surrounding facilities that produce synthetic fuel is inevitable. The main areas of concern are:

1. Surface Disturbance
2. Water Management
3. Atmospheric Emissions
4. Service Network

The oil sand operation is based on the mining of bitumen bearing sand. The disturbance of the terrain has a considerable effect on the wildlife of the area and the potential for returning the area to its natural state after the duration of the project.

Water management covers the withdrawal and discharge of water from natural sources as well as changes made to the natural watershed during the project life.

Atmospheric emissions are the gaseous by-products that are released to the atmosphere. The effects of these effluents have far reacting consequences and often necessitate additional process development for the removal of toxic constituents.

The service network covers the changes that are made to allow a facility to function properly. These may include the construction of highways, pipelines as well as any of the other facilities required.

The oil sand facilities are all located in the Fort McMurray area. This environment is quite delicate and the environmental problems are severe.

3.8.2 Surface Disturbances

The three main sources of surface disturbance are:

the open pit mine;
the plant site;
the tailings ponds.

The mining procedure is described in Section 3.2.

The open pit mine is the largest area that will be affected. The Alsands project envisages an area of 26.8 km^2 being used as a mine. The muskeg removed from the mine site will be stored for future revegetation. The overburden that covers the oil sand will be removed and used to build up starter dykes around the tailings ponds. Once the tailings ponds are complete the overburden will be used to fill the mine pits.

The muskeg that is removed is stored to permit water drainage into the surrounding muskeg.

The Fort McMurray area is characterized by an extensive artesian water system. These water sources must be depressurized during the mining process.

The tailings ponds affect large areas of land. Alsands predict that approximately 21 km^2 (and a depth of 50 m) will be used for tailings ponds, while Syncrude uses approximately 28 km^2 of land (and a depth of 42 m). Apart from the adverse environmental effects, the large ponds prevent recovery of bitumen deposits from the land on which they are situated.

Alsands predict that after 8 years of operation sufficient oil sand will have been mined for the tailings to be returned to the mine site. The sludge that is produced is stored in the tailings pond (mined out area) where the sand particles settle. The sludge that remains in suspension is pumped into the starter dyke where subsequent settling takes place. The sand filled area is covered with topsoil and revegetated. In the starter dyke, a clarified water layer forms over a suspended solids layer after approximately 3 years. Any subsequent settling takes decades. The sludge is quite toxic (pH=8.2) and is a definite environmental concern. While there is no proven method of land reclamation at the moment, a considerable amount of research examining this problem is being performed. The environmental problems caused by the Hot Water Extraction Process necessitate a solution to the tailings problem.

The plant site and the associated area (which include the camp, plant site, storage area and service network) required for plant operation are a similar size to that required by the mine or tailings pond. The Alsands project anticipates using some 21 km^2 of land for these facilities.

The oil sand plants have a project life estimated at 25 years. When the projects are completed, a land reclamation scheme will also be finalized. The main area that cannot be reclaimed with today's technology is the portion of the starter pond that contains unsettled sludge. The mine pit that is filled with tailings sand and overburden as well as the plant site are covered with the stored muskeg and then revegetated. The Suncor facility has had reasonable success with revegetation of their original mine area after being initially hampered by small rodents eating the seedlings. This type of land reclamation has been fairly successful.

3.8.3 Water Management

The watershed in the Fort McMurray area is an important component of the environment. The ground is covered with a layer of muskeg which contains a large precentage of water. The muskeg is an extremely delicate form of vegetation that reduces wind and water erosion. The water contained within the surface material that is removed is drained and fed into the surrounding rivers. The removal of the muskeg reduces the ability of the rivers to absorb sudden changes in flow. Storage ponds are provided to minimize the change during precipitation.

The Hot Water Extraction Process uses large volumes of water in the bitumen extraction step. The water used is contained within the tailings sludge that is stored in the tailings ponds. The initial settling stage takes approximately three years before any of the water can be recycled to the extraction plant. The water that is required in the process is withdrawn from the local rivers, normally the Athabasca. The amount that is withdrawn for the first three years is approximately three times the steady state amount shown in the material balance (Section 3.1). The water requirements drop to the steady state requirement after three years when the tailings pond clarified water layer develops, providing a recycle water source.

The Athabasca River is the major water source in the Fort McMurray area. It is estimated that the proposed Alsands plant at maximum consumption (ie. zero recycle) will require about 3% of the minimum water flow. The cumulative consumption of the oil sands facilities will have an affect on the localized river environment. This effect can be minimized if additional synthetic crude facilties are started at staggered intervals greater than three years. This would allow each facility to develop its own recycle water supply and would therefore reduce the load placed on the Athabasca River.

The artesian water system is another source that must be managed to ensure smooth operation of the mine. This system contains water at fairly high pressures which must be relieved before a full scale mining operation can proceed. The water that is obtained from the aquifers is processed through settling ponds and filtration systems before being discharged into local rivers.

Water contained in the tailings sludge is another potential groundwater problem. A certain amount of water drains through the dyke walls. To prevent dyke failure, filter drains are built into the dyke wall. The drains convey the water to ditches surrounding the dyke, where it is pumped back into the centre of the tailings pond. As the tailings pond develops, the fine particles seal the dyke wall and reduce the seepage.

3.8.4 Atmospheric Emissions

The oil sand facilities produce three main significant atmospheric emissions, sulphur compounds, particles and water vapour.

The bitumen that is extracted from the oil sand has a high sulphur content 0.05 mass fraction. It is treated in a fluid coker where a fuel gas, a burner gas, an LPG product, a hydrocarbon stream and a coke stream are produced. The coke is gasified to produce hydrogen with hydrogen sulphide and slag as byproducts. The hydrogen produced is combined with the hydrocarbon stream in hydrotreaters to produce the synthetic crude product and a fuel gas. The fuel gas streams produced in the fluid coker and the hydrotreaters are treated in conventional amine units. The sour gas produced by the amine unit is combined with the hydrogen sulphide stream from the gasifier as feed to a Claus plant for sulphur removal. The Claus plant tail gas is then treated in a SCOT tail gas cleanup unit. The Claus-SCOT plant combination has an estimated 99.9% sulphur recovery. The tail gas from the SCOT plant is combined with the burner gas from the fluid coker in a CO boiler where any hydrogen sulphide is oxidized to sulphur dioxide. This stream is then released to the atmosphere through a stack.

This sulphur is disposed of through four main streams, the fuel gas (which contains approximately 72% of the sulphur fed to the fluid coker), the hydrogen from the gasifier (18%), the burner gas (7%) and the synthetic crude product (3%). The other product steams contain trace amounts of sulphur but these have no significant effect on the environment. The hydrogen sulphide that is produced in the gasifier and the fuel gas contains 90% of the sulphur that was introduced within the bitumen. The Claus-SCOT plant reduces this to less than 1% and produces a solid sulphur product that corresponds to effectively 90% of the sulphur fed to the process.

The burner gas is not suitable for processing by the Claus-SCOT plant. It is composed of nitrogen, carbon monoxide and a small percentage of sulphur compounds. This stream is fed directly to the CO boiler where any hydrogen sulphide is oxidized to sulphur dioxide and carbon monoxide is converted to carbon dioxide. The majority of the sulphur that is vented to the atmosphere originates with the burner gas. This stream is quite large so that even small percentages of sulphur give an overall sulphur output that is large.

The Alberta Government has set the ground level sulphur dioxide concentration at 0.2×10^{-6} volume fraction for a ½ hour interval (20) as well as recommended emission standard of 3.2 tonnes of sulphur per 1000 m^3 of bitumen fed to the upgrading plant. The Alsands facility will meet the ground level concentration standard but will fail to meet the standard of sulphur per unit of bitumen. Several methods of desulphurization have been considered for CO boiler flue gas treatment. These systems generally involve contacting a cooled flue gas with an absorbent and then reheating the flue gas for disposal into the atmosphere. These schemes have not been proven on this type of service and as the capital cost is in the range of $100 million, there is a reluctance to apply this type of technology. Another problem with this type of flue gas treatment is the production of a lime based sludge byproduct. Enormous volumes of this byproduct would have to be contained and an environmentally satisfactory disposal program would require development. Alsands has decided that this type of flue gas treatment is unjustified and do not plan to use this technology. The ERCB has conditionally approved this but has requested that the plant be constructed in such a way that this system could be added at a later date. The existing plants, Suncor and Syncrude, have no such provision.

The particulate that is produced by the oil sands facilities is mainly composed of ash and coke fines. These are produced in the fluid coker and go through the process with the burner gas to the CO boiler. An electrostatic precipitator that is located downstream from the CO boiler reduces the particulate concentration to within standards set by the government agencies. If more stringent standards are set, the electrostatic precipitators could be used in conjunction with bag houses.

The water vapour problem is only apparent during cold weather. The tailings that are produced by the Hot Water Extraction Process enter the tailings ponds at approximately 70°C. The effluent liquid produces clouds of steam that can produce localized icy conditions and reduced visibility. At present, there is no economic solution to this problem. The only recourse is to store the tailings away from the transportation network and reduce the hazard potential by conventional means (for example - sanding the roads).

3.8.5 Service Network

The development of a 19000 m^3/sd of synthetic crude facility would require accommodation for approximately 10,000 people. Fort McMurray accommodates the staff for Syncrude and Suncor.Studies that were performed for the Alsands proposal found the option of building new towns instead of expanding Fort McMurray to be perferrable. As the oil sands cover such a large area, it is likely that the new town would be used automatically to reduce the transportation time between the plant site and the support town.

The construction of a new town would necessitate the implementation of all support services. Fort McMurray is serviced by railroad, highway and airline. Electricity and natural gas are also available. These services would have to be extended to any new town as well as pipelines to transport the products.

The corridor concept will reduce the environmental impact of implementing the service network. This concept consists of a narrow corridor that contains power lines, highways and buried pipelines. The minimization of area affected reduces the environmental impact to a minimum.

3.8.6 Conclusion

The technology that is being used in the new oil sands facilities reduces the environmental impact considerably but a few areas need to be developed so that the bitumen can be removed in a responsible manner. A tailings treatment method or a new extraction process is required so that the tailings do not cover land and render it non reclaimable. A practical means of flue gas desulphurization is also required. This would reduce any sulphur emissions to a level that will comply with forseeable legislation until the year 2000.

The service network is a problem that cannot be avoided if the oil sands are to be developed. The corridor concept will keep this environmental impact to a minimum.

The existing oil sands facilities are renowned for sporadic operation and plant upsets. These sudden changes in plant operation make the effluent streams extremely difficult to manage. While these sudden spillages and minor discharges are undesirable, they are also unavoidable. With today's technology and as experience develops, smoother operation and reduced environmental impact should be possible.

STREAM NUMBER	MOL WT	S.G.	°API	101	102	103	104	110	111	201	202	203	204	205	206	207	208	209	210
STREAM	COMPONENT PROPERTIES			FROM PRIMARY EXTRACTION BITUMEN RECOVERY	REJECT FROM EXTRACTION RECOVERY	TAILINGS FROM EXTRACTION	BLENDED TAILINGS TO TAILINGS POND	DILUTED BITUMEN FROM FROTH TREATMENT	TAILINGS FROM FROTH TREATMENT	WATER TO SOUR WATER SYSTEM	DILUENT RECYCLE	BITUMEN FROM DILUENT RECOVERY UNIT	NAPHTHA AND WATER TO FROTH TREATMENT	BUTANE & NAPHTHA TO NAPHTHA	FUEL GAS TO AMINE PLANT	COKE FINES FROM CO	FLUE GAS TO STACK FROM ESP	FLUE GAS FROM ESP AND HEAT RECOVERY	COKE PRODUCT FROM FLUID COKER
COMPONENT				m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h
BITUMEN				933	3	58	58	911	22			21 720							
WATER				446	34	8 007	8 007	88	393		288		12				97.4	183.0	101
SOLIDS				97	210	6 549	6 549	9	88										
NAPHTHA								593	13		18 562		606	4 740					
DISTILLATE																			
BUTANE														190					
FUEL GAS															50.3				
CO															15.3				
H_2S															25.2				
COKE																4.0			86
N_2 + Ar																	634.6	2 042.2	
SO_2																	6.2	6.3	
CO_2																	184.2	292.1	
O_2																	5.6	274.1	
WATER (SOUR)										910									
TOTALS				1 476	247	14 614	14 614	1 601	516	910	18 850	21 720	618	4 930	90.8	4.0	918	2 805	187
t/d				35 424	5 928	350 736	350 736	38 424	12 384	910	14 518	21 864	14 832	3 744	2 178	96	22 032	67 320	4 488
m³/d								42 404		910	18 850	21 720	19 260	4 930					
SP GR										~1									
kg/m³										1000									
Sm³/d																	30 751 722	68 250 756	
TJ/d																			
Note 1				4	4	4	4	4	4	2	2	2	4	2	2	2	2	2	2

REMARKS

BASIS FOR FLOWS -

STREAM 205 - ASSUMED EQUAL TO MASS FLOW OUT OF NAPHTHA HYDROPROCESSOR

Note 1: No. of parallel streams for total flows above

STP conditions: 60° F, 14.7 PSIA; m³ conditions 15°C, 101 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'KD	APP'D
A	82-06-4	ORIGINAL	SHNL		

COMPONENT MATERIAL BALANCE

Client: EMR

Plant: LIQUID FUELS FROM OIL SANDS

Project No: 1058

DYNAWEST PROJECTS LTD

TABLE 3.1

Sheet 1 of 10

FORM NO 11-01-10 81

STREAM	COMPONENT PROPERTIES				211 Distillate to hydro-treating from fluid coker	212 Gas oil to hydro-treating from fluid coker	213 Butane & naphtha from fluid coker	214 Gas from fluid coker	215 Burner gas to CO boiler from fluid coker	216 [illegible]	217 LPG to amine plant	221 Flue gas from CO boiler to ESP	222 HP steam from CO boiler	223 Sour water [illegible]	224 Sour water from [illegible]	225 Sour water [illegible]	226 Sour water [illegible]	227 Gas to sulfur plant [illegible]	228 HP steam to utilities plant	229 [illegible] pond
COMPONENT	MOL WT	S.G.	°API		m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h
WATER (SOUR)										Δ				37	2	29	31	15		
COKE									4.0			4.0								
BITUMEN																				
BUTANE							137													
NAPHTHA							3 423													
DISTILLATE					4 800															
GAS OIL						10 330														
WATER/STEAM												97.4	275.5							
N_2												629.6							360	9 597
SO_2								Δ	6.11			6.2								
H_2S								30			7.8							8		
CO								20	97.1		4.7									
FUEL GAS								58.3												
LPG											15.5									
COS								Δ												
CO_2												184.2								
O_2												6.6								
TOTALS					4 800	10 330	3 560	108.3	107.2	Δ	28.0	923	275.5	37	2	29	31	23	360	9 597
t/d					4 224	10032	2 704	2 599	2 573	Δ	672	22 128	6 612	888	48	696	744	552	8 640	230 328
m³/d					4 800	10 530	3 560													230 328
SP GR					0.88	0.97														
kg/m³																				
Sm³/d												30731772								
TJ/d													22.9						29.89	
Note 1					2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	1

REMARKS

BASIS FOR FLOWS:-

Δ: COMPONENTS IDENTIFIED BUT INDIVIDUAL FLOWS UNKNOWN

Note 1: No. of parallel streams for total flows above

STP conditions: 60°F, 14.7 PSIA; m³ conditions: 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
0	82-8-6	ORIGINAL	DWL		

COMPONENT MATERIAL BALANCE

Client: EMR

Plant: LIQUID FUELS FROM OIL SANDS

Project No. 1053

DYNAWEST PROJECTS LTD.

TABLE 3.1

Sheet 2 of 10

FORM NO. M-01-1081

NUMBER STREAM COMPONENT	MOL WT	SG	°API	294 m³/d [✓] t/h	295 m³/d [✓] t/h	296 m³/d [✓] t/h	297 m³/d t/h [✓]	298 m³/d [✓] t/h	301 m³/d [✓] t/h	302 m³/d [✓] t/h	303 m³/d [✓] t/h	304 GAS FROM NAPHTHA HYDRO-PROCESSOR m³/d t/h [✓]	305 m³/d t/h [✓]	306 GAS FEED TO GAS RECOVERY UNIT m³/d t/h [✓]	307 SYNTHETIC CRUDE TO CRUDE STORAGE m³/d [✓] t/h	311 m³/d [✓] t/h	312 m³/d t/h [✓]	313 m³/d [✓] t/h	314 m³/d [✓] t/h
WATER				12 168	295	8 760	164	3 910											
BUTANE									720						720			53	190
NAPHTHA										3 930	410				3 930			1 317	4 740
DISTILLATE															4 570	4 570			
GAS OIL															9 830	9 830			
SO_2												Δ	Δ	Δ			Δ		
H_2S												0.84	3	33			2.16		
CO												Δ	Δ	20			Δ		
FUEL GAS/LPG												2.1	7.5	65.8			5.3		
COS												Δ	Δ	Δ			Δ		
TOTALS				12 168	295	8 760	164	3 910	720	3 930	410	2.94	10.5	118.8	19 050	14 400	7.46	1 370	4 930
t/d				12 168	295	8 760	3 936	3 910	417	3 013	314	70.6	252	2 050	16 091	12 661	~179	1 040	3 740
m³/d				12 168	295	8 760	3 936	3 910	720	3 930	410				19 050	14 400		1 370	4 930
SP GR									0.58	0.766	0.766								
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1				2	2	2	1	2	2	2	1	2	2	2	2	2	2	2	2

REMARKS

BASIS FOR FLOWS:-

Δ COMPONENTS IDENTIFIED BUT INDIVIDUAL FLOWS UNKNOWN

Note 1: No. of parallel streams for total flows above

SCF conditions: 60° F, 14.7 PSIA, m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
A	82-04-06	ORIGINAL			

COMPONENT MATERIAL BALANCE

Client EMR
Plant LIQUID FUELS FROM OIL SANDS
Project No. 1058

DYNAWEST PROJECTS LTD.

TABLE 3.1
Sheet 3 of 10

FORM NO. M-01-1081

NUMBER / STREAM / COMPONENT	MOL WT	SG	°API	401 HYDROGEN PLANT PRODUCT HYDROGEN [✓ m³/d] t/h	402 HYDROGEN TO DISTILLATE (GAS OIL) HYDRO-PROCESSING [✓ m³/d] t/h	403 HYDROGEN TO NAPHTHA HYDRO-PROCESSING [✓ m³/d] t/h	404 SLAG DISPOSAL FROM HYDROGEN PLANT m³/d [✓ t/h]	405 HYDROGEN SULPHIDE TO SULPHUR RECOVERY FROM HYDROGEN PLANT m³/d [✓ t/h]	406 HYDROGEN PLANT BLOWDOWN [✓ m³/d] t/h	407 CO₂ VENT FROM HYDROGEN PLANT m³/d [✓ t/h]	408 [illegible] PLANT m³/d [✓ t/h]
HYDROGEN				4 170 000	3550000	620 000					
ASH							9.48				
WATER							9.64		344		19
H_2S								7.36			
S							0.17				
COS								1.07			
CO_2								16		228	
O_2											
Ar/N_2											
FUEL GAS											
LPG											
TOTALS				4 170 000	3550000	620 000	19.29	24.43	344	228	19
t/d				353	301	52	463	586	344	5472	456
m³/d									344		
SP GR											
kg/m³											
Sm³/d				4 170 000	3 550 000	620 000					
TJ/d											
Note 1				4	2	2	4	4	4	4	4

NUMBER / STREAM / COMPONENT	501 OXYGEN TO HYDROGEN PLANT m³/d [✓ t/h]	502 HP STEAM FROM OXYGEN PLANT m³/d [✓ t/h]	503 N_2 FROM OXYGEN PLANT m³/d [✓ t/h]	601 FUEL GAS FROM [illegible] PLANT [✓ m³/d] t/h	602 PLANT FUEL TO UTILITIES [✓ m³/d] t/h	603 PLANT FUEL TO HEAT [illegible] BOILER [✓ m³/d] t/h	604 PLANT FUEL TO GAS TURBINES [✓ m³/d] t/h	605 PLANT FUEL TO PROCESSING SECTION [✓ m³/d] t/h
HYDROGEN								
ASH								
WATER		65.3						
H_2S								
S								
COS								
CO_2								
O_2	62.7		7.4					
Ar/N_2	1.6		231.3					
FUEL GAS				1 780 565	1 621 557	326 657	1 299 700	159 148
LPG								
TOTALS	64.4	65.3	238.7	1 780 565	1 621 357	326 657	1 299 700	159 148
t/d	1545	1568	5729	1208	1050	220	879	109
m³/d								
SP GR								
kg/m³								
Sm³/d				1 780 565	1 621 557	326 657	1 299 700	159 148
TJ/d		5.4		58.0	52.8	10.6	42.2	5.2
Note 1	2	2	2	2	2	2	2	2

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SCF conditions: 60° F, 14.7 PSIA; m³ conditions 15°C, 101 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
A	82-09-[illegible]	ORIGINAL	[illegible]		

COMPONENT MATERIAL BALANCE

Client EMR

Plant LIQUID FUELS FROM OIL SANDS

Project No. 1058

DYNAWEST PROJECTS LTD.

TABLE 3.1

Sheet 4 of 10

FORM NO. M-01-1081

NUMBER	COMPONENT PROPERTIES			606	607	611	621	701	702	703	704	705	706	711	712	713	714	721	722
STREAM				LPG TO FUEL OR SALES	GAS TO SULPHUR PLANT FROM [illegible]	SULPHUR PRODUCT TO SULPHUR STORAGE	TAIL GAS TO CO BOILER	WATER SUPPLY FROM SETTLING BASIN	RECYCLE MAKE UP TO [illegible]	WATER TO WATER CLARIFIER AND STORAGE	RAW WATER TO WATER STORAGE	WATER TO CLARIFIER	WATER TO CLARIFIER	RECYCLE WATER OUTPUT	COLD WATER INJECTION TO [illegible]	RECYCLE WATER TO UTILITIES AREA	RECYCLE WATER TO [illegible] COOLING	WATER FROM OIL SEPARATOR	SOLIDS FROM OIL SEPARATORS
COMPONENT	MOL WT	S.G.	°API	[✓ m³/d] t/h	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d [✓ t/h]	[✓ m³/d] t/h	m³/d [✓ t/h]	m³/d [✓ t/h]	[✓ m³/d] t/h	m³/d [✓ t/h]
FUEL GAS																			
LPG				711															
WATER							45	70 031	6 492	63 539	≤63 539	≤63 539	63 539	6 867	0	6 867	10 392	4 824	
											Σ=63 539								
SOLIDS																			
														77		77	77		9
S						38.9													
CO																			
SO_2					20		20												
H_2S					33		0.044												
COS																			
CO_2																			
N_2+Ar							186												
TOTALS				711	53	38.9	251	70 031	6 492	63 539	≤63 539	≤63 539	63 539	6 944	0	6 944	10 469	4 824	9
t/c				371	1 272	933	6024	70 031	6 492	63 539	<63 539	≤63 539	63 539	166 647	0	166 647	251 256	4 824	216
m³/d				711	—			70 031	6 492	63 539	≤63 539	≤63 539	63 539	164 800	0	164 800	249 412	4 824	—
SP GR																			
kg/m³									—										
Sm³/d																			
TJ/d				17.0															
Note 1				3	2	2	2	1	1	1	1	1	1	1	4	2	2	2	2

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

STD conditions - 60° F, 14.7 PSIA, m³ conditions 15°C, at 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
A	82-04-[illegible]	ORIGINAL	DWE		

COMPONENT MATERIAL BALANCE

Client: EMR

Plant: LIQUID FUELS FROM OIL SANDS

Project No. 1058

DYNAWEST PROJECTS LTD

TABLE 3.1

Sheet 5 of 10

FORM NO. M-01-1081

NUMBER	COMPONENT PROPERTIES			801	802	803	804	810	820	821	822	823	824	825	826	827	828	829	830
STREAM				OIL SAND FROM MINE AREA	OIL SAND FROM SURGE STOCKPILE	[illegible]	WATER [illegible] AT COKE [illegible]	COKE [illegible] TO [illegible] PLANT	[illegible]	WATER TO OIL [illegible]	[illegible]	RAIN TO TAILINGS POND	[illegible]	[illegible]	[illegible]	[illegible]	AIR TO [illegible]	AIR TO [illegible] & FLUID COKE	AIR TO OXYGEN PLANT
COMPONENT	MOL WT	SG	°API	m³/d [✓ t/h]	m³/d [✓ t/h]	[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d [✓ t/h]	[✓ m³/d] t/h	m³/d [✓ t/h]	[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d [✓ t/h]	m³/d [✓ t/h]	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d [✓ t/h]	m³/d [✓ t/h]
BITUMEN				994	994				21 720				911						
WATER				443	443	13 205	2 434			38	563 537	25 819	50	6 232	32 581	70 031			
SOLIDS				6 779	6 779			-		9			0	77					
NAPHTHA													593						
FLUID COKE								90.21											
NITROGEN/AR																	27 573 [illegible]	419	233
OXYGEN																	6 904 864	123	70
TOTALS				8 216	8 216	13 205	2 434	90.21	21 720	47	563 537	25 819	1 554	6 309	32 581	70 031	34 27[illegible]	542	303
t/d				197 196	197 196	13 205	2 434	2 165	21 864	1 128	563 557	25 819	37 296	151 416	32 581	70 031	1 891	13 009	7 271
m³/d						13 205	2 434		21 720	914	563 537	25 819	41 490	149 560	32 581	70 051			
SP GR																			
kg/m³																			
Sm³/d																	34 379 [illegible]		
TJ/d																			
Note 1				4	4	4	2	1	2	2	1	1	2	1	1	1	2	2	2

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

ST conditions - 60° F, 14.7 PSIA; m³ conditions 15°C, 101 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
A	82-09-06	ORIGINAL	DWK		

COMPONENT MATERIAL BALANCE

Client EMR

Plant LIQUID FUELS FROM OIL SANDS

Project No 1058

DYNAWEST PROJECTS LTD.

TABLE 3.1

Sheet 6 of 10

FORM NO M-01-1081

Component	MOL WT	SG	°API	831 AIR TO FLUID COKER m³/d t/h	832 AIR TO CO BOILER m³/d t/h	899 FLUE GAS FROM STACK TO ATMOS. m³/d t/h	901 GAS FROM [illegible] TO HEAT RECOVERY m³/d t/h	902 FLUE GAS [illegible] HEAT RECOVERY BOILER m³/d t/h
NITROGEN				186	438	675	28 275 995	28 [illegible]
ARGON							350 945	348 312
OXYGEN				53	125	31	5 044 556	4 781 546
CO_2						93	1 108 733	1 471 649
SO_2						2	145	1 117
WATER						60	2 110 168	2 670 430
TOTALS				239	563	927	37 255 000	37 255 000
t/d				5 755	13 509	22 248	45 072	45 288
m³/d								
SP GR								
kg/m³								
Sm³/d						22 557 351	37 255 000	37 266 000
TJ/d								
Note 1				2	2	1	2	2

Component	903 HP STEAM FROM HEAT RECOVERY BOILER m³/d t/h	904 HP STEAM TO [illegible] COMPRESSOR TURBINES m³/d t/h	905 CONDENSATE [illegible] m³/d t/h	906 LP STEAM TO CONDENSING TURBINE m³/d t/h	907 HP STEAM TO BACK PRESSURE TURBINES m³/d t/h	908 STEAM BLOW DOWN FROM STEAM [illegible] m³/d t/h	909 [illegible] m³/d t/h	911 LP STEAM TO [illegible] DRIVING m³/d t/h	912 LP STEAM TO [illegible] FROM UTILITIES m³/d t/h	913 LP STEAM TO HEAT EXCHANGERS m³/d t/h	914 STEAM TO [illegible] PLANT m³/d t/h
NITROGEN											
ARGON											
OXYGEN											
CO_2											
SO_2											
WATER	369	709	0	64	646	20	434	646	646	≤ 113	≥ 535 ≤ 646
TOTALS	369	709	0	64	646	20	430	646	696	≤ 113	≥ 557 ≤ 696
t/d	8 855	17 035	0	1 530	16 505	480	10 416	15 505	15 506	≤ 2 709	≥ 12 73[illegible] ≤ 15 696
m³/d											
SP GR											
kg/m³											
Sm³/d											
TJ/d	30.7	59.0	0.0	5.3	53.7	1.6		53.6	43.6		
Note 1	2	2	2	1	2	2	2	2	2	2	2

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SCF conditions -60° F, 14.7 PSIA; m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
A	82-09-06	ORIGINAL	[illegible]		

COMPONENT MATERIAL BALANCE

Client EMR

Plant: LIQUID FUELS FROM OIL SANDS

Project No 1050

DYNAWEST PROJECTS LTD.

TABLE 3.1

Sheet 7 of 10

FORM NO. H-01-1081

STREAM NUMBER	MOL WT	SG	[illegible]	916	916	917	918	913	920	921	922	923	924	931	932	933	934	935	936
STREAM				LP STEAM TO DEAERATOR	LP STEAM [illegible]	LP STEAM TO [illegible] RECOVERY	NET STEAM TO [illegible] RECOVERY	STEAM [illegible]	L.P. STEAM TO [illegible]	L.P. STEAM TO FLUID COKER	L.P. STEAM [illegible] (HEATING)	LP STEAM TO FLUID COKER	LP STEAM TO [illegible] PLANT	HOT WATER FROM [illegible] COOLER	[illegible] TANK	HOT WATER TO [illegible] PLANT	HOT WATER TO [illegible]	[illegible]	[illegible]
COMPONENT				m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	(m³/d) t/h	m³/d (t/h)	m³/d (t/h)
H_2O				≤ 113	533	378	344	34	23	132	Δ	Δ	32	10 392	10392	10 392	84 613	7 700	< 113
SOLIDS														77	77	77	TRACE	77	
TOTALS																			
t/h				≤ 113	533	378	344	34	23	132	Δ	Δ	32	10 469	10 469	10 469	84 613	7 777	≤ 113
m³/d				≤ 2 709	12 796	9 072	8 256	816	562	3 172	Δ	Δ	2 218	251 256	251 252	251 256	84 613	186 648	≤ 2 709
SP GR														243 412	243 412	243 412	84 613	184 800	
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1				2	2	4	4	4	4	2	2	2	2	2	2	2	2	4	2

REMARKS

BASIS FOR FLOWS:-

Δ : COMPONENTS IDENTIFIED BUT INDIVIDUAL FLOWS UNKNOWN

Note 1: No. of parallel streams for total flows above

SCF conditions - 60° F, 14.7 PSIA; m³ conditions 15°C, 101 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
A	[illegible]	ORIGINAL	DWL		

COMPONENT MATERIAL BALANCE

Client EMR

Plant LIQUID FUELS FROM OIL SANDS

Project No. 1058

DYNAWEST PROJECTS LTD.

TABLE 3.1

Sheet 8 of 10

FORM NO. 11-01-1081

NUMBER				937	938	939	940	941	942	943	944	945	946	947	951	952	953	954	955
STREAM	COMPONENT PROPERTIES			L.P. STEAM FROM [illegible]	CONDENSATE FROM CONDENSING TURBINE	CONDENSATE TO DEAERATOR	COOLING WATER FROM TOWER	COOLING WATER [illegible] TO [illegible]	WATER CLARIFIED TO WATER FILTRATION	WATER FROM [illegible] COOLING TOWER	GLAND WATER & SEAL WATER FILTRATION	POTABLE & GLAND WATER TO EXTRACTION	POTABLE & GLAND WATER TO UPGRADING AREA	WATER TO OFFSITES AREA	BFW TO EXTRACTION	BFW TO UPGRADING AREA CO BOILER	BFW TO UTILITIES PLANT	BFW TO [illegible] SECTION	BFW TO HEAT RECOVERY BOILERS
COMPONENT	MOL WT	SG	°API	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d [✓ t/h]	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d t/h	[✓ m³/d] t/h
H_2O			–	0	69	69	1926	2 426	63 539	11993	26 591	20 001	6 590	6 867	18 794	8 363	8 855	0	8 855
SOLIDS														77					
TOTALS				0	69	69	1226	2 426	63 539	11 993	26591	20 001	6590	6 344	18 794	8 363	8 855	0	8 855
t/d					1 530	1 530	46 231	58 224	63 639	11993	26 591	20001	6 590	166 697	18 794	8 863	8 856	0	8 855
m³/d					1 530	1 530	46 231	58 224	63 637	11993	26 521	20001	6 530	169 759	18 794	8 363	8 855	0	8 855
SP GR																			
kg/m³																			
Sm³/d																			
TJ/d				0													5.4	0.0	5.4
Note 1				2	1	1	1	1	1	1	1	2	2	2	2	2	2	2	2

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SCF conditions - 60°F, 14.7 PSIA; m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
0	82-09-06	ORIGINAL			

COMPONENT MATERIAL BALANCE

Client EMR

Plant LIQUID FUELS FROM OIL SANDS

Project No. 1050

DYNAWEST PROJECTS LTD.

TABLE 3.1

Sheet 9 of 10

FORM NO. H-01-1081

NUMBER STREAM COMPONENT	MOL WT	S.G.	°API	956 TREATED WATER TO REACTION	957 TREATED WATER & CONDENSATE	958 BFW TO HYDROGEN PLANT	959 BFW TO CO BOILER	960 BFW TO [illegible] PLANT	961 BFW TO OXYGEN PLANT	995 TOTAL [illegible] FROM UTILITIES	996 UTILITIES BLOWDOWN	998 WATER [illegible] COOLING [illegible]	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h
				[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d [✓ t/h]	[✓ m³/d] t/h	[✓ m³/d] t/h	[✓ m³/d] t/h	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]							
H_2O				24 955	26 485	73	6 612	9 939	1 576	481	454	27							
TOTALS				24 955	26 485	73	6 612	9 939	1 576	481	454	27							
t/d				24 965	26 485	1 751	6 612	9 939	1576	11 544	10 896	648							
m³/d				24 955	26 485	1 751	6 612	9 939	1576	11 544	10 896	648							
SP GR																			
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1				1	1	4	2	2	2	2	2	2							

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above
STP conditions: 60° F, 14.7 PSIA, m³ conditions 15°C, KN 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
A		ORIGINAL			

COMPONENT MATERIAL BALANCE
Client EMR
Plant LIQUID FUELS FROM OIL SANDS
Project No. 1050
DYNAWEST PROJECTS LTD.
TABLE 3.1
Sheet 10 of 10

FORM NO. H-01-1081

Table 3.6.1 Oil Sands Process Major Utilities Summary

	Power (MW)		Steam (t/sd)		Recycle Water (GJ/h)(3)	
	Produced	Consumed	Produced	Consumed	Produced	Consumed
Mine		49.4				
Extraction		48.6		12 325[1]		2 433
Primary Upgrading		30.9		5 067		
Secondary Upgrading		84.5	21 145	4 427	2 707	
Emission & Gas Clean Up		2.5		5 012		
Oxygen Plant		0.1	1 568			
Utilities & Offsites	240.3	24.3	13 147	9 029		
TOTAL	240.3	240.3	35 860	35 860	2 707	2 433[2]

NOTE:

1. Includes Steam to Extraction Exchangers.
2. Low grade heat that can not be utilized is rejected to atmosphere.
3. Recycle water is used as cooling water.

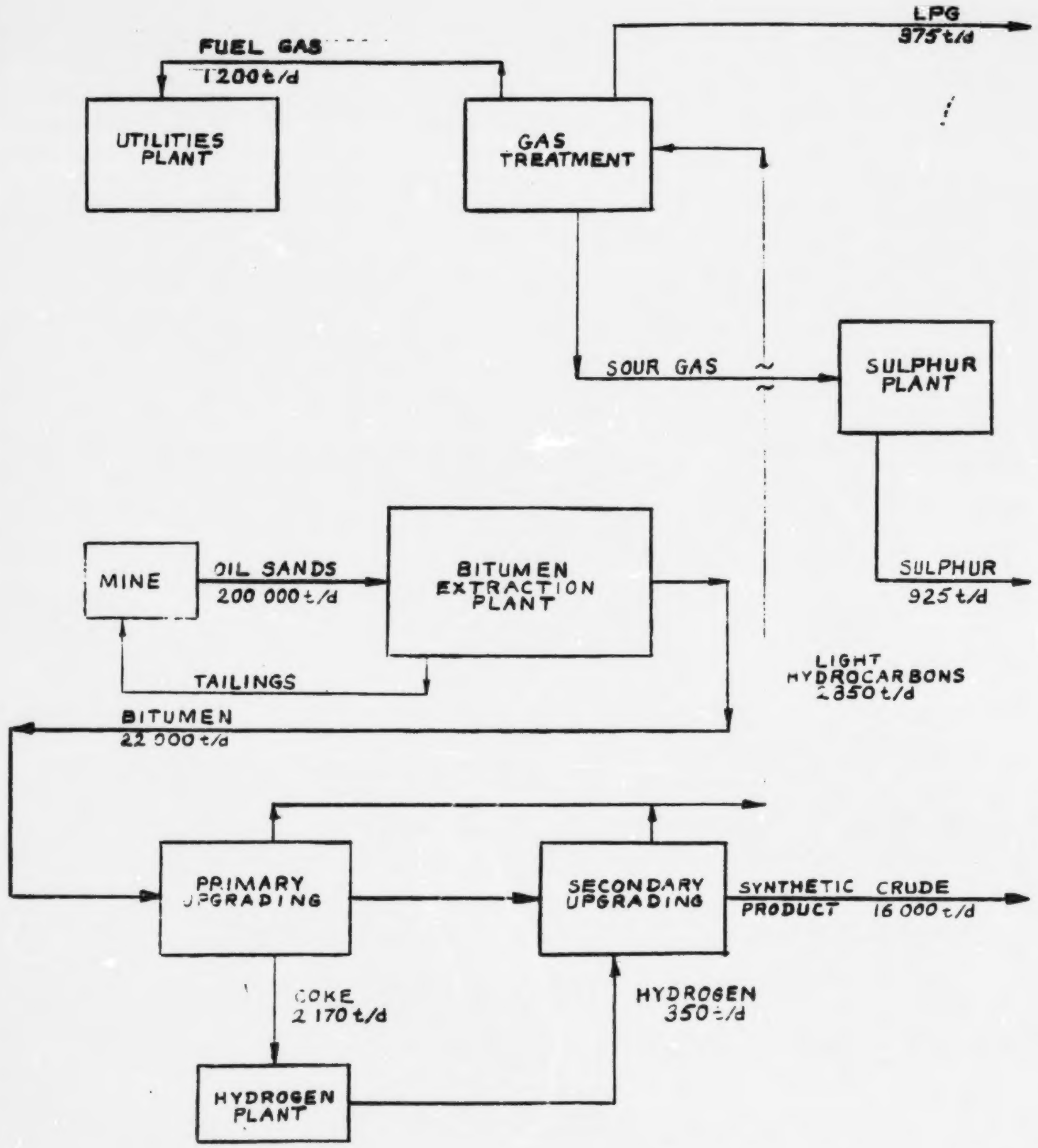
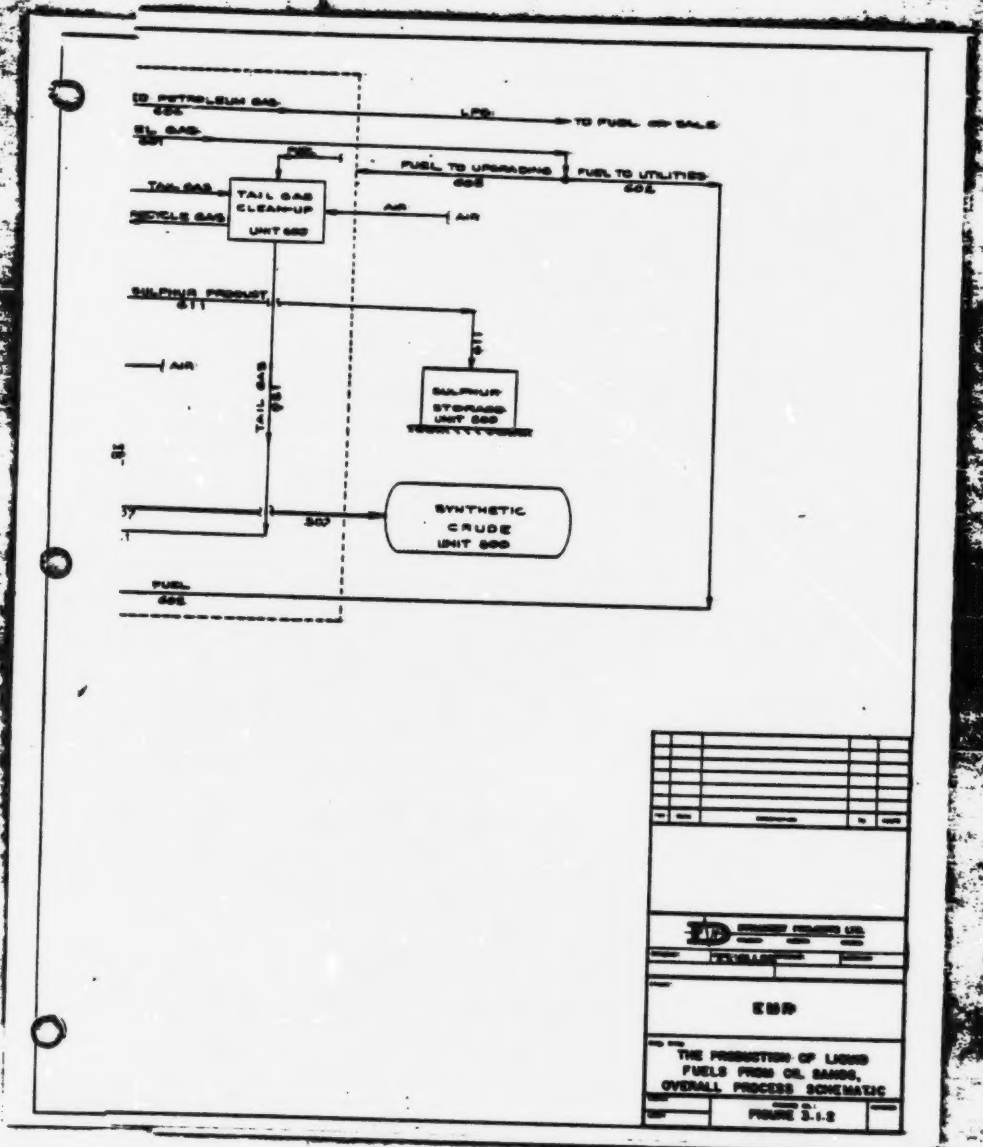

FUEL GAS
1200 t/d
LPG
375 t/d
UTILITIES PLANT
GAS TREATMENT
SOUR GAS
SULPHUR PLANT
MINE
OIL SANDS
200 000 t/d
BITUMEN EXTRACTION PLANT
SULPHUR
925 t/d
TAILINGS
LIGHT HYDROCARBONS
2850 t/d
BITUMEN
22 000 t/d
PRIMARY UPGRADING
SECONDARY UPGRADING
SYNTHETIC CRUDE PRODUCT
16 000 t/d
COKE
2 170 t/d
HYDROGEN
350 t/d
HYDROGEN PLANT

FIGURE 3.1.1
LIQUID FUELS FROM OIL SANDS
OVERALL BLOCK DIAGRAM

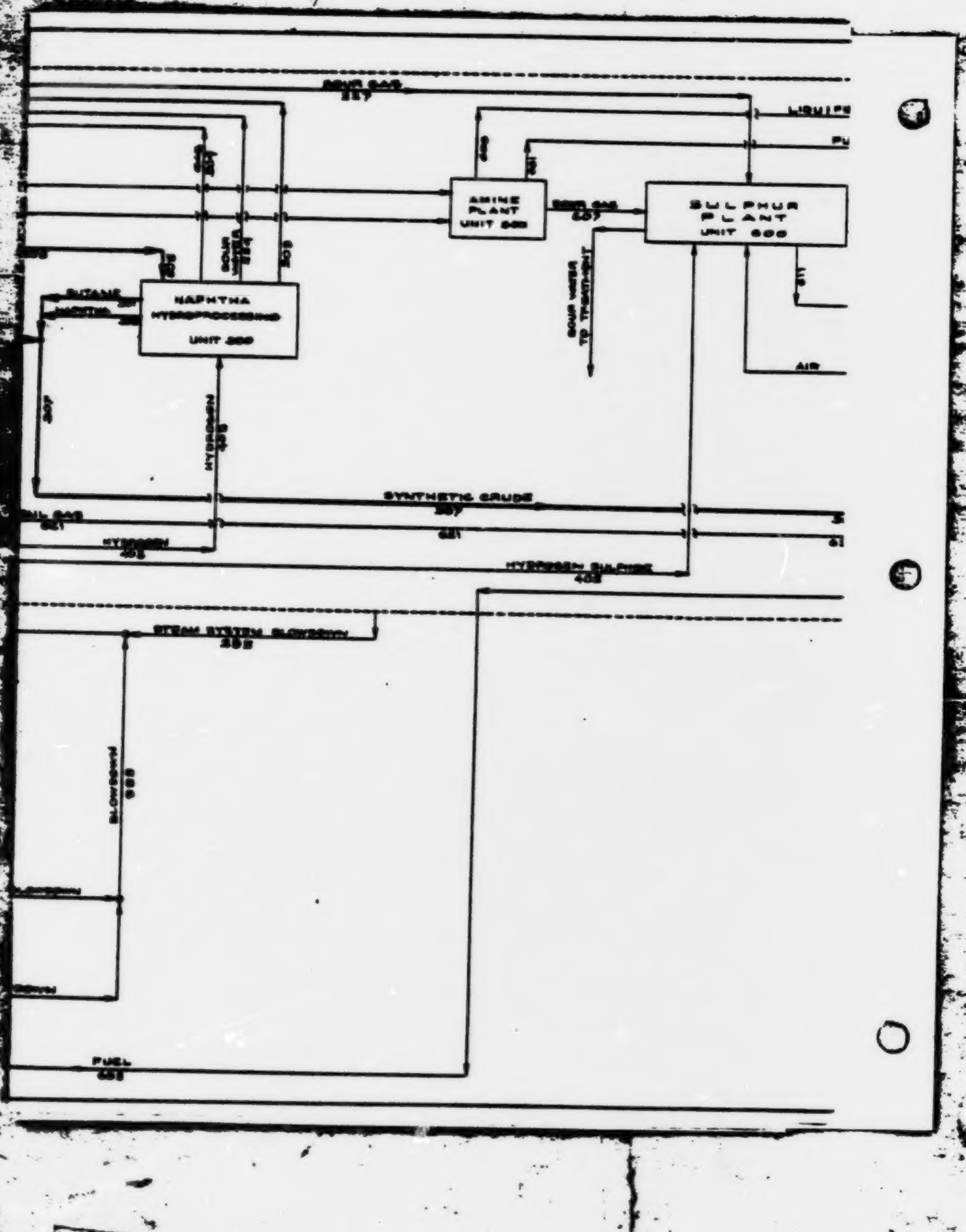
DYNAWEST PROJECTS LTD.

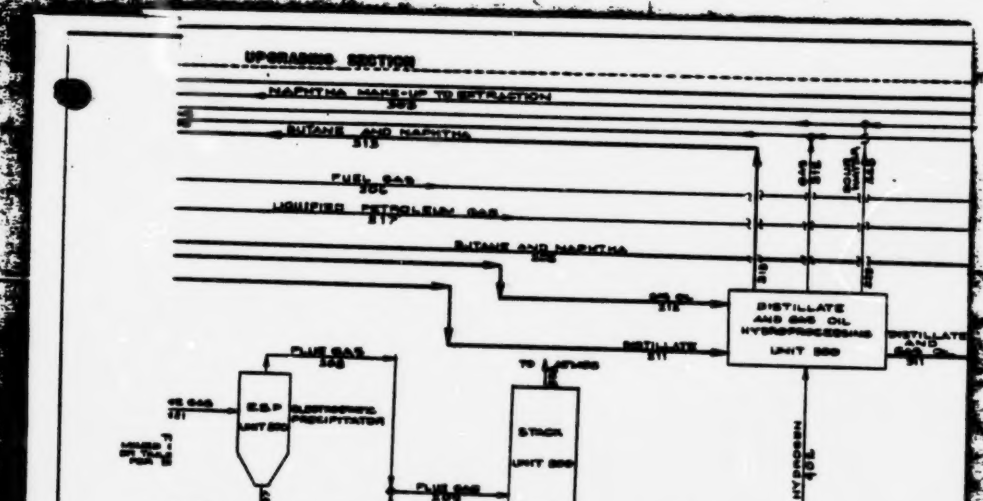
ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET
						CONT SHT

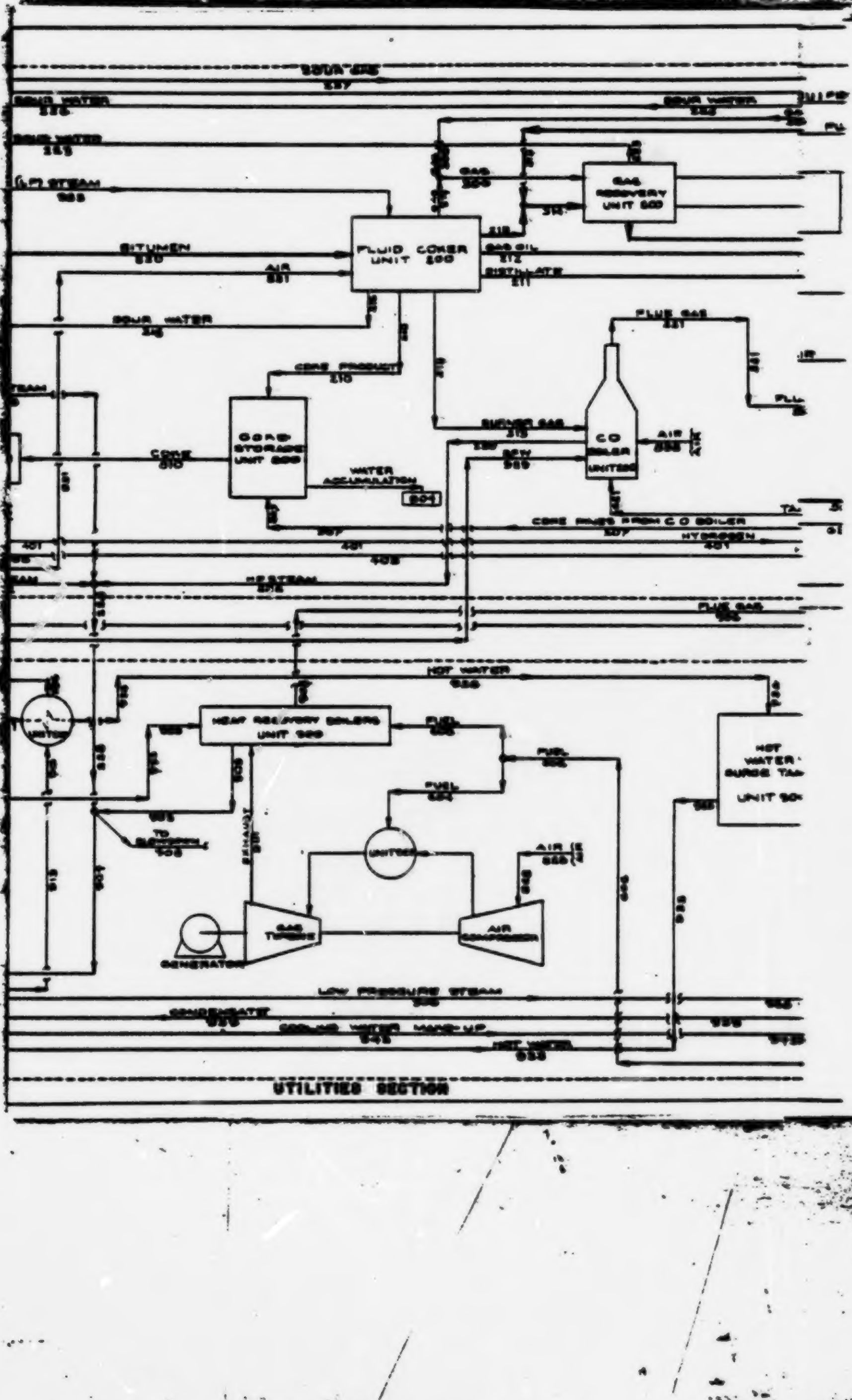

LPG
TO FUEL OR SALE
FUEL
FUEL TO UPGRADING
FUEL TO UTILITIES
TAIL GAS
RECYCLE GAS
TAIL GAS CLEAN-UP UNIT
AIR
AIR
SULPHUR PRODUCT
611
611
AIR
TAIL GAS
SULPHUR STORAGE UNIT
SYNTHETIC CRUDE UNIT
307
FUEL
EMR

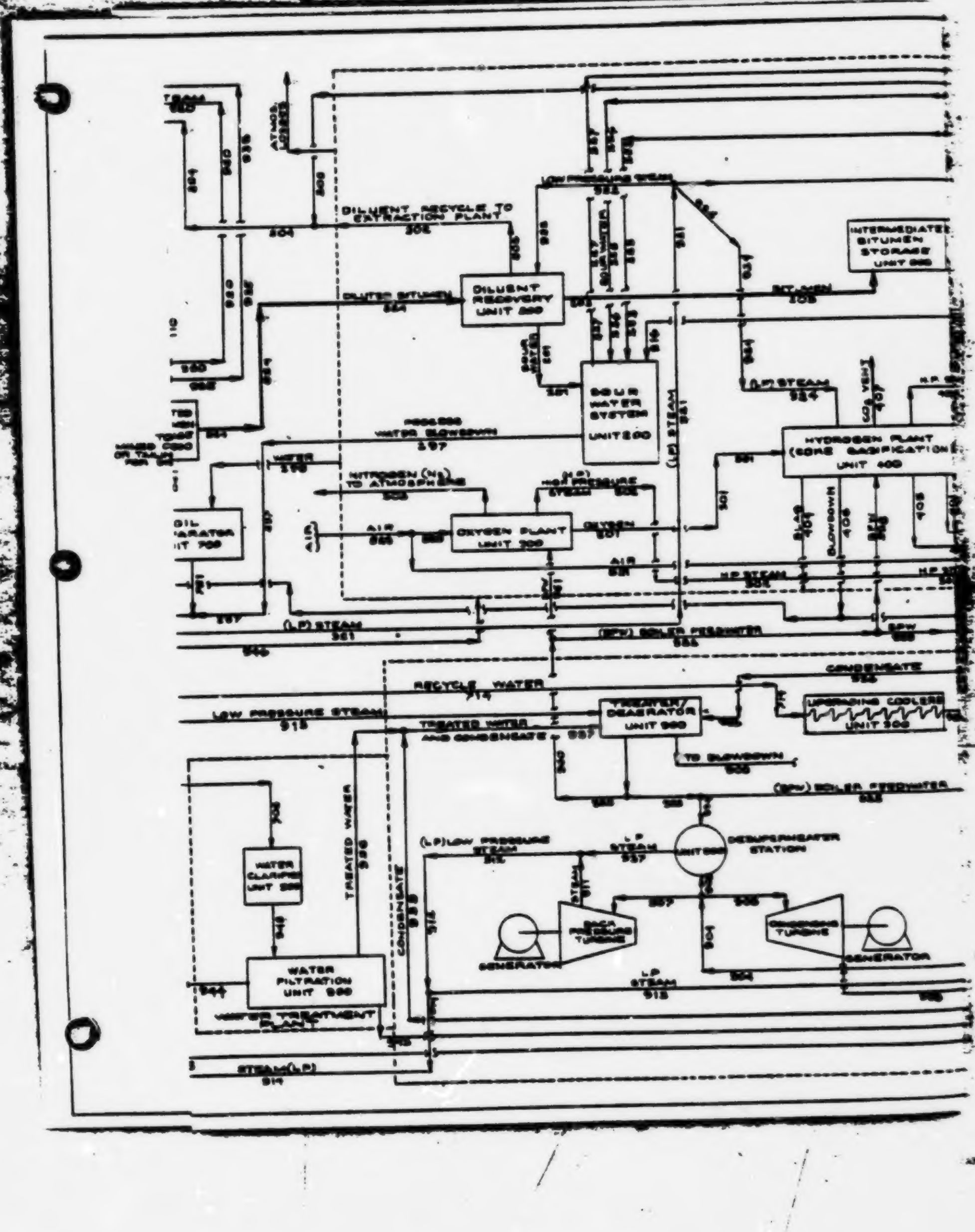
THE PRODUCTION OF LIQUID FUELS FROM OIL SANDS, OVERALL PROCESS SCHEMATIC

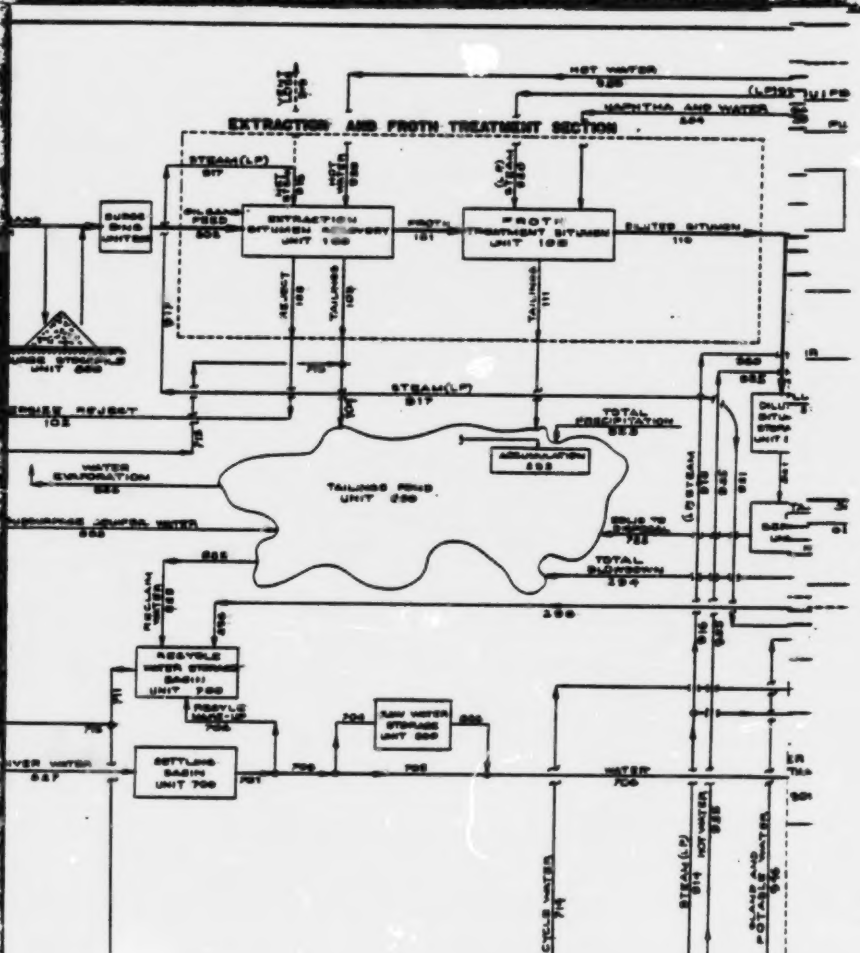
FIGURE 3.1.2

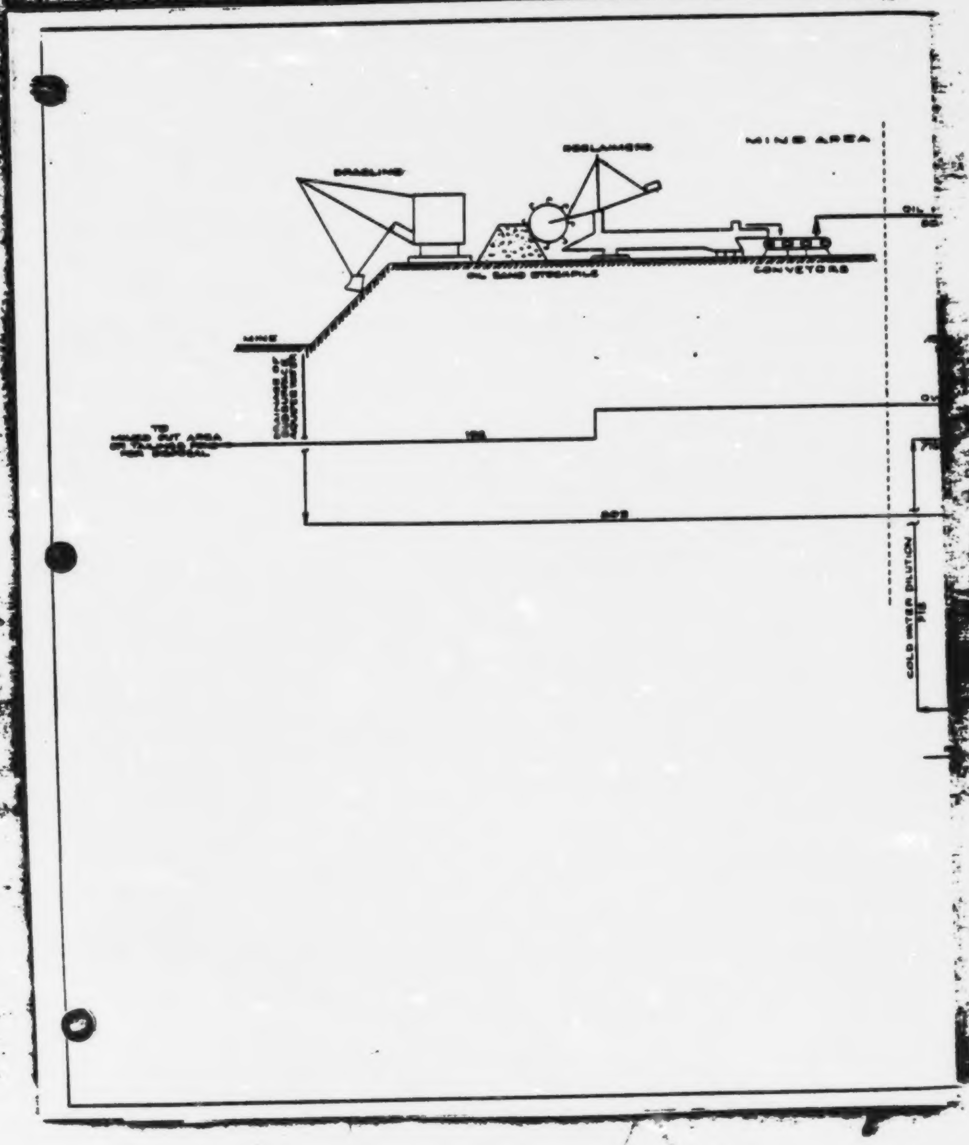
SOUR GAS
LIQUIFI
FU
AMINE PLANT UNIT 600
SOUR GAS
SULPHUR PLANT UNIT 600
SOUR WATER TO TREATMENT
AIR
BUTANE
NAPHTHA
NAPHTHA HYDROPROCESSING UNIT 300
GAS
SOUR WATER
HYDROGEN
307
SYNTHETIC CRUDE
TAIL GAS
HYDROGEN
HYDROGEN SULPHIDE
STEAM SYSTEM BLOWDOWN
BLOWDOWN
FUEL

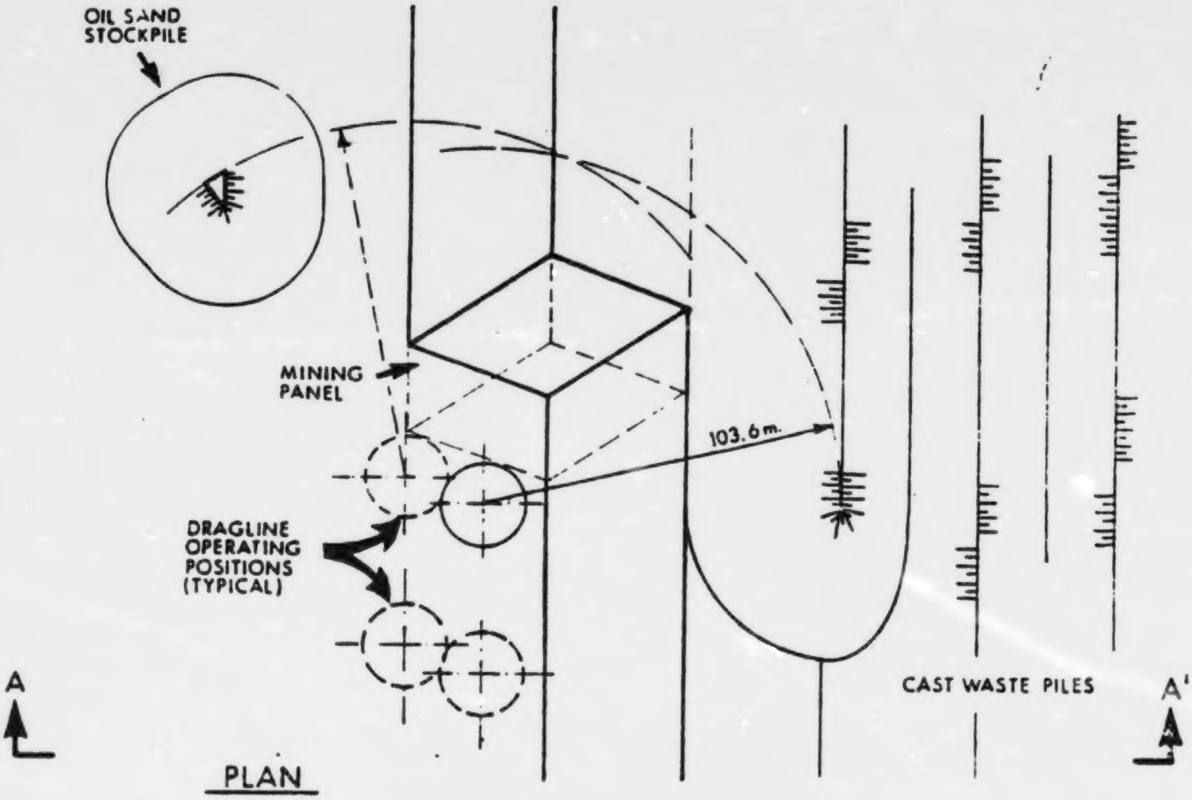

UPGRADING SECTION
NAPHTHA MAKE-UP TO EXTRACTION
BUTANE AND NAPHTHA
313
FUEL GAS
LIQUIFIED PETROLEUM GAS
217
BUTANE AND NAPHTHA
GAS OIL
212
DISTILLATE
211
DISTILLATE AND GAS OIL HYDROPROCESSING
UNIT 300
DISTILLATE AND GAS OIL
FLUE GAS
E.S.P
UNIT 800
ELECTROSTATIC PRECIPITATOR
TO ATMOSPHERE
STACK
UNIT 800
FLUE GAS
HYDROGEN
TAIL GAS
HYDROGEN SULPHIDE
HYDROGEN
COOLING TOWER BLOWDOWN
COOLING WATER TOWER
UNIT 900
UNIT 900
WATER
WATER ATMOSPHERIC LOSS
COOLING WATER TOWER
UTILITIES' BLOW


SOUR GAS
SOUR WATER
SOUR WATER
(LP) STEAM
GAS
GAS RECOVERY UNIT 600
FLUID COKER UNIT 200
BITUMEN
AIR
GAS OIL
212
DISTILLATE
211
SOUR WATER
FLUE GAS
COKE PRODUCT
210
COKE STORAGE UNIT 800
COKE
810
WATER ACCUMULATION
804
SUPPER GAS
CO BOILER
AIR
BFW
COKE FINES FROM C.O BOILER
HYDROGEN
401
401
HP STEAM
FLUE GAS
HOT WATER
HEAT RECOVERY BOILERS UNIT 900
FUEL
FUEL
FUEL
TO
EXHAUST
AIR
GAS TURBINE
AIR COMPRESSOR
GENERATOR
HOT WATER SURGE TANK UNIT 90
LOW PRESSURE STEAM
CONDENSATE
COOLING WATER MAKE-UP
HOT WATER
UTILITIES SECTION

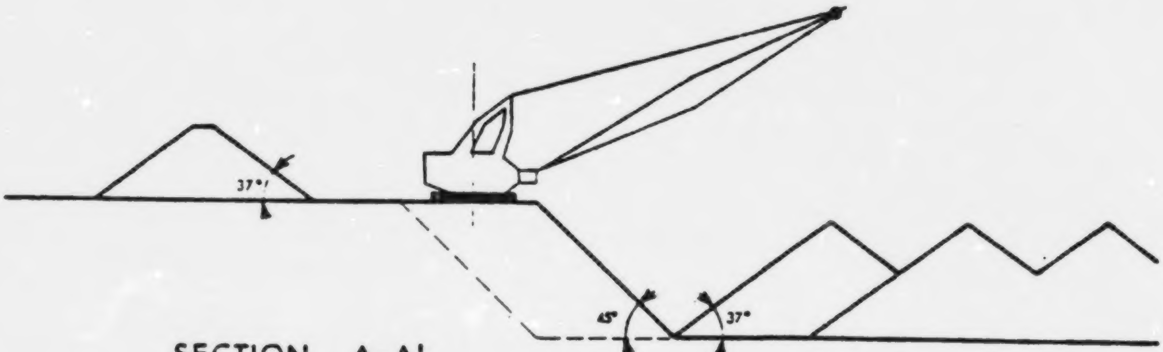

STEAM
ATMOS. LOSSES
DILUENT RECYCLE TO EXTRACTION PLANT
302
DILUTED BITUMEN
DILUENT RECOVERY UNIT 200
LOW PRESSURE STEAM
INTERMEDIATE BITUMEN STORAGE
BITUMEN
205
SOUR WATER
SOUR WATER SYSTEM UNIT 200
(LP) STEAM
PROCESS WATER BLOWDOWN
297
WATER
NITROGEN (N2) TO ATMOSPHERE
(HP) HIGH PRESSURE STEAM
AIR
OXYGEN PLANT UNIT 300
OXYGEN
301
OIL SEPARATOR UNIT 700
HYDROGEN PLANT (COKE GASIFICATION UNIT 400
CO2 VENT
407
SLAG
401
BLOWDOWN
406
405
H.P. STEAM
(LP) STEAM
(BFW) BOILER FEEDWATER
BFW
CONDENSATE
RECYCLE WATER
714
LOW PRESSURE STEAM
TREATED WATER AND CONDENSATE
TREATER/DEAERATOR UNIT 900
UPGRADING COOLERS UNIT 300
TO BLOWDOWN
(BFW) BOILER FEEDWATER
WATER CLARIFIER UNIT 900
TREATED WATER
CONDENSATE
(LP) LOW PRESSURE STEAM
LP STEAM
DESUPERHEATER STATION
UNIT 900
STEAM
BACK PRESSURE TURBINE
CONDENSING TURBINE
GENERATOR
GENERATOR
WATER FILTRATION UNIT 900
944
WATER TREATMENT PLANT
LP STEAM
STEAM (LP)
914


EXTRACTION AND FROTH TREATMENT SECTION
HOT WATER
NAPHTHA AND WATER
STEAM (LP)
917
SURGE BINS
OIL SAND FEED
EXTRACTION BITUMEN RECOVERY UNIT 100
FROTH
101
FROTH TREATMENT BITUMEN UNIT 105
DILUTED BITUMEN
110
REJECT
TAILINGS
TAILINGS
111
STEAM (LP)
917
TOTAL PRECIPITATION
ACCUMULATION
WATER EVAPORATION
TAILINGS POND UNIT 290
TOTAL BLOWDOWN
294
RECLAIM WATER
RECYCLE WATER STORAGE BASIN UNIT 700
RECYCLE MAKE-UP
RAW WATER STORAGE UNIT 800
RIVER WATER
SETTLING BASIN UNIT 700
701
WATER
RECYCLE WATER
STEAM (LP)
914
HOT WATER
GLAND AND POTABLE WATER
RECYCLE WATER
715


MINE AREA
DRAGLINE
RECLAIMERS
OIL SAND STOCKPILE
CONVEYORS
OIL
MINE
TO MINED OUT AREA OR TAILINGS POND FOR DISPOSAL
COLD WATER DILUTION


OIL SAND
STOCKPILE
MINING
PANEL
103.6 m.
DRAGLINE
OPERATING
POSITIONS
(TYPICAL)
A
CAST WASTE PILES
A'

PLAN



SECTION A-A'

FIGURE 3.2.1

Taken from reference (2)

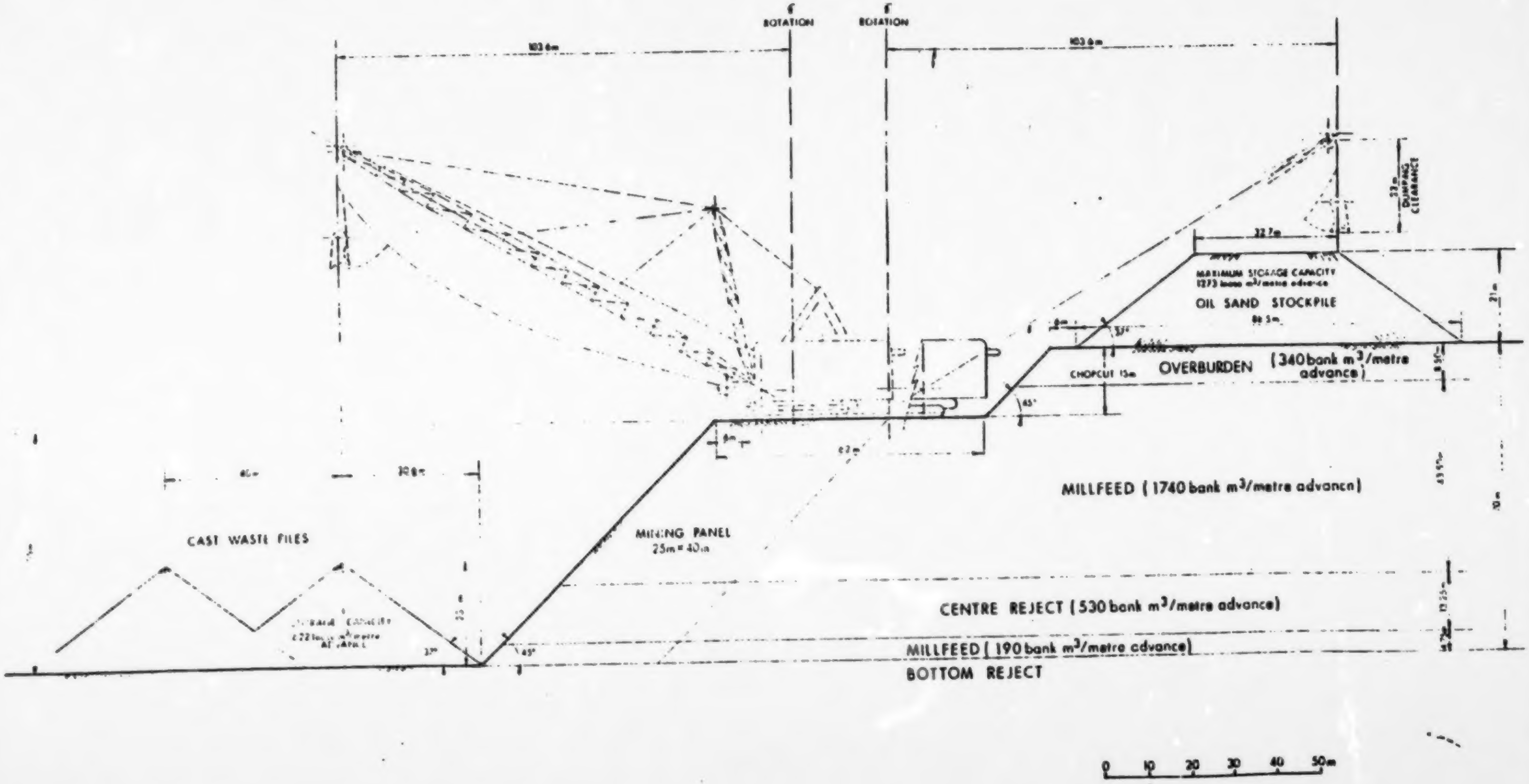

℄ ROTATION
℄ ROTATION
103.6m
103.6m
22m DUMPING CLEARANCE
32.7m
MAXIMUM STORAGE CAPACITY
1273 loose m³/metre advance
OIL SAND STOCKPILE
86.5m
21m
37°
CHOPCUT 15m
OVERBURDEN (340 bank m³/metre advance)
45°
62m
MILLFEED (1740 bank m³/metre advance)
MINING PANEL
25m × 40m
CAST WASTE PILES
46m
30.8m
37°
45°
CENTRE REJECT (530 bank m³/metre advance)
MILLFEED (190 bank m³/metre advance)
BOTTOM REJECT
0
10
20
30
40
50m

FIGURE 3.2.2

Taken from reference (4)

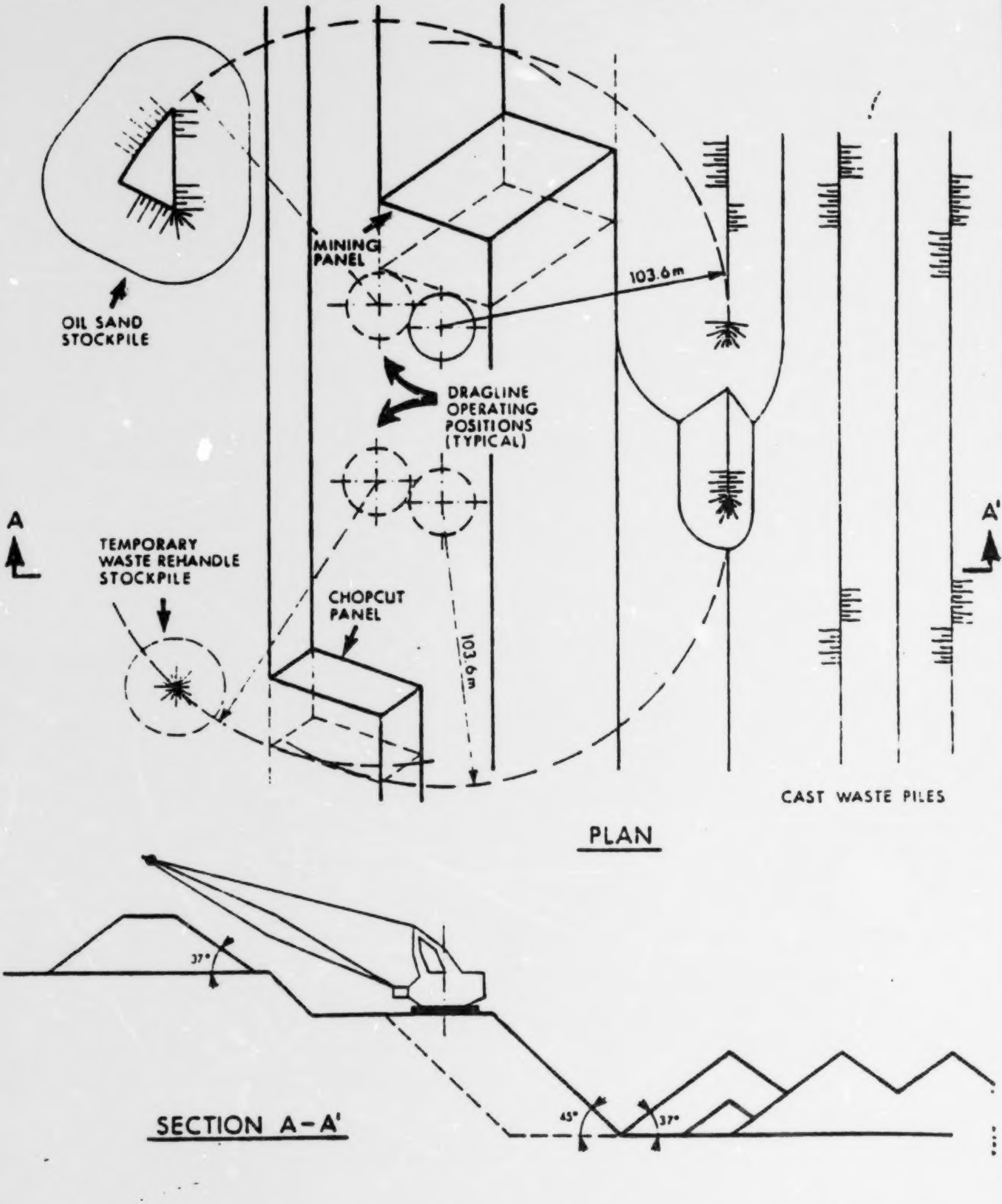
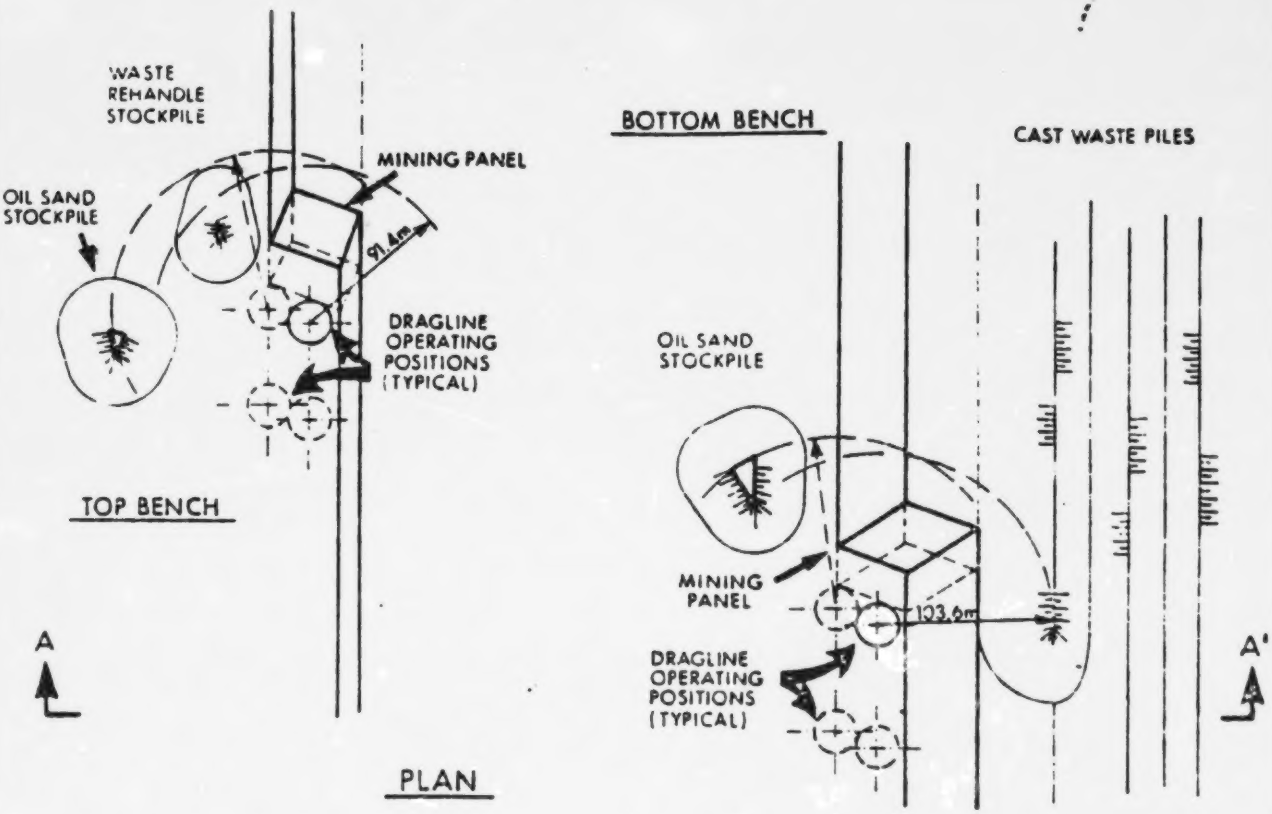
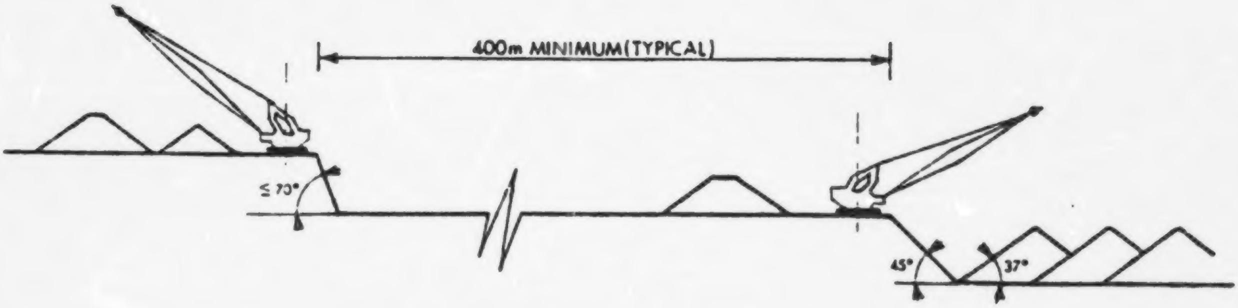

OIL SAND STOCKPILE
MINING PANEL
103.6m
DRAGLINE OPERATING POSITIONS (TYPICAL)
A
A'
TEMPORARY WASTE REHANDLE STOCKPILE
CHOPCUT PANEL
103.6m
CAST WASTE PILES
PLAN
37°
SECTION A-A'
45°
37°

FIGURE 3.2.3

Taken from reference (2)


WASTE REHANDLE STOCKPILE
OIL SAND STOCKPILE
MINING PANEL
91.4m
DRAGLINE OPERATING POSITIONS (TYPICAL)
TOP BENCH
A
BOTTOM BENCH
CAST WASTE PILES
OIL SAND STOCKPILE
MINING PANEL
103.6m
DRAGLINE OPERATING POSITIONS (TYPICAL)
A'

PLAN


400m MINIMUM (TYPICAL)
≤ 70°
45°
37°

SECTION A-A'

FIGURE 3.2.4

Taken from reference (2)

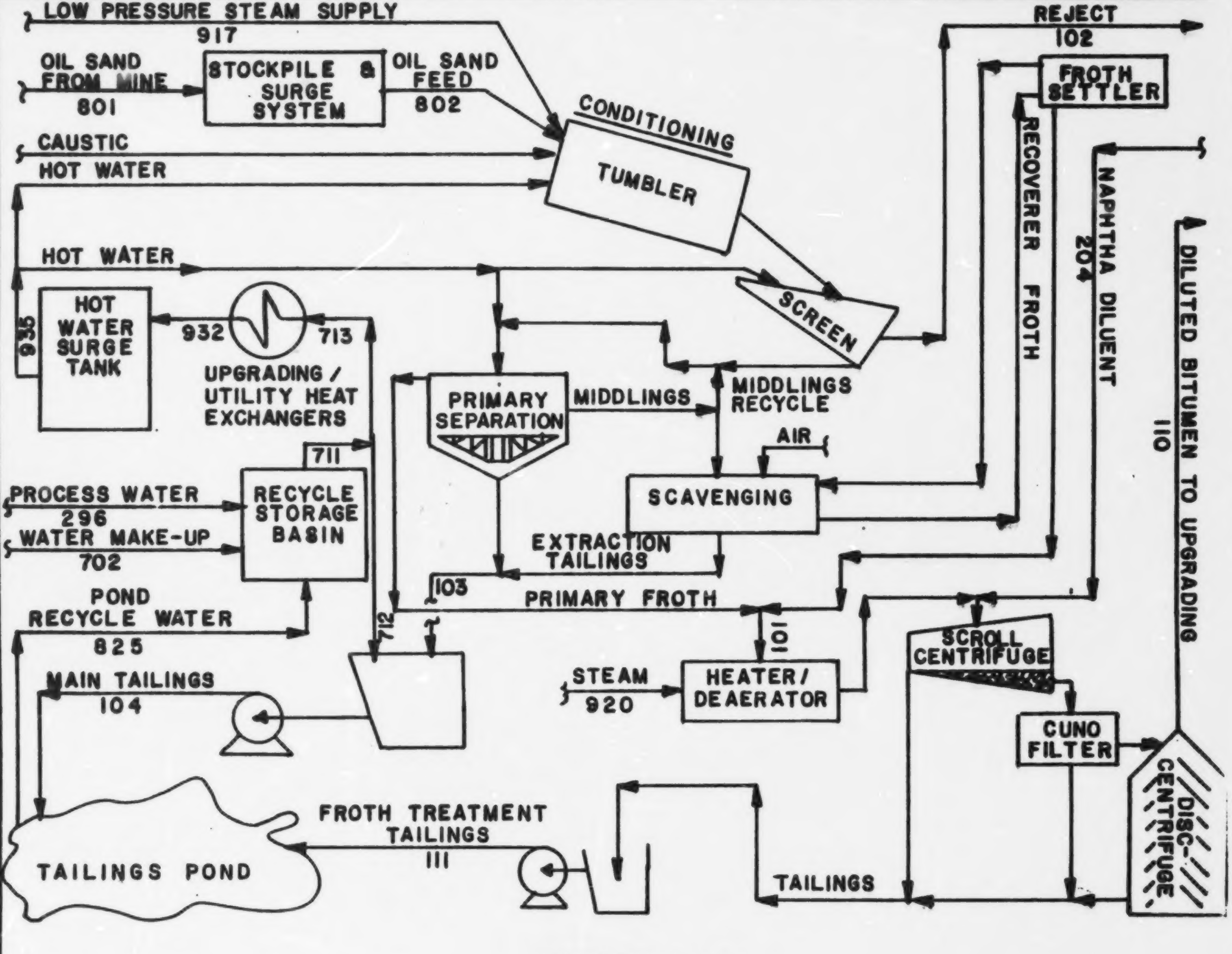

LOW PRESSURE STEAM SUPPLY
917
OIL SAND FROM MINE
801
STOCKPILE & SURGE SYSTEM
OIL SAND FEED
802
CONDITIONING
CAUSTIC
HOT WATER
TUMBLER
HOT WATER
SCREEN
REJECT
102
FROTH SETTLER
RECOVERER FROTH
NAPHTHA DILUENT
204
DILUTED BITUMEN TO UPGRADING
110
935
HOT WATER SURGE TANK
932
713
UPGRADING / UTILITY HEAT EXCHANGERS
711
PRIMARY SEPARATION
MIDDLINGS
MIDDLINGS RECYCLE
AIR
SCAVENGING
EXTRACTION TAILINGS
103
PRIMARY FROTH
101
PROCESS WATER
296
WATER MAKE-UP
702
RECYCLE STORAGE BASIN
POND RECYCLE WATER
825
712
MAIN TAILINGS
104
STEAM
920
HEATER/ DEAERATOR
SCROLL CENTRIFUGE
CUNO FILTER
DISC-CENTRIFUGE
FROTH TREATMENT TAILINGS
111
TAILINGS POND
TAILINGS

FIGURE 3.3.1
FLOW SCHEMATIC OF EXTRACTION & FROTH TREATMENT PLANTS

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

SHEET
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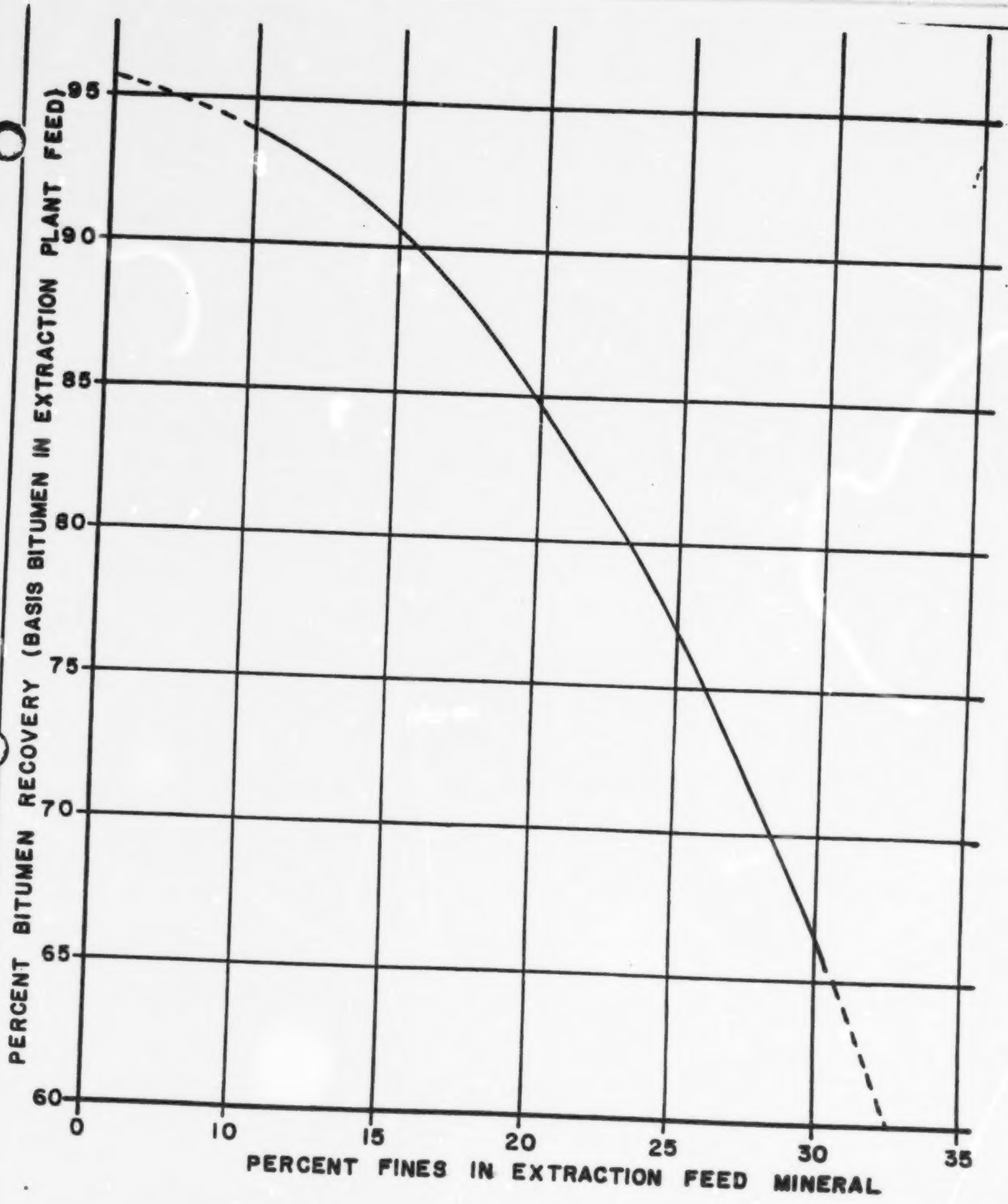
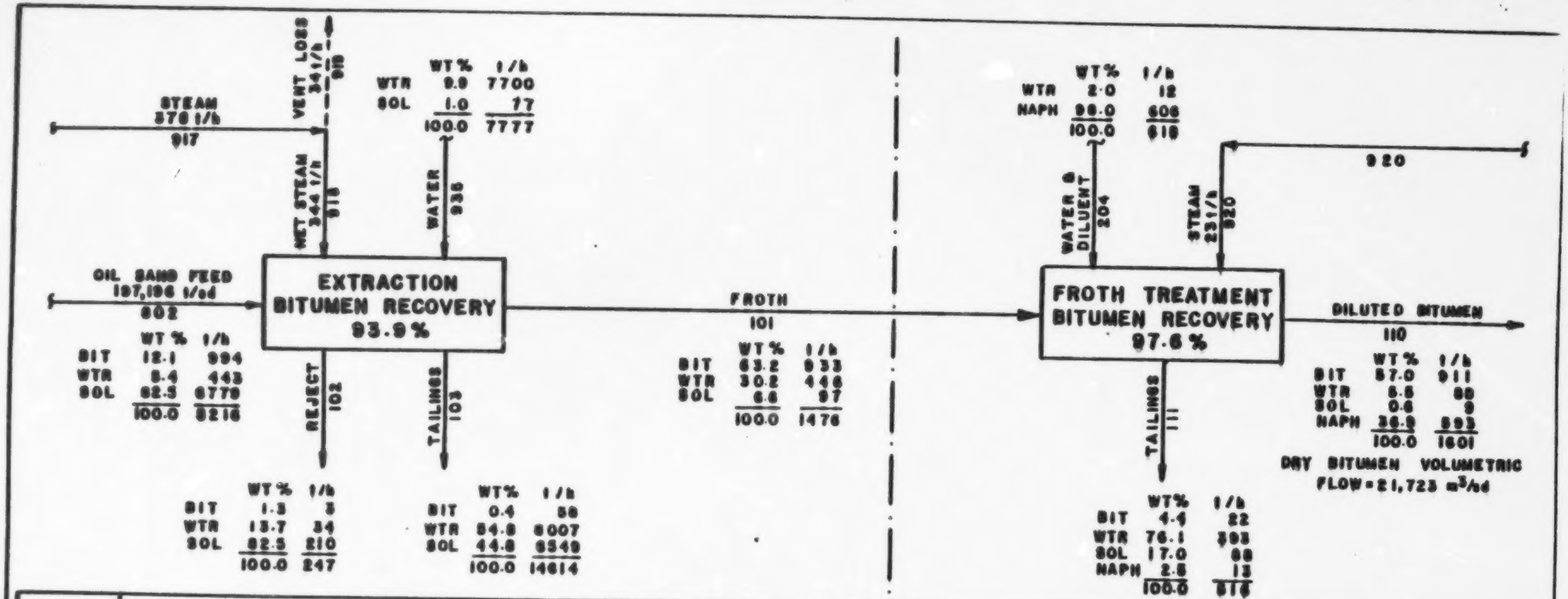


FIGURE 3.3.2
BITUMEN RECOVERY VERSUS FINES IN EXTRACTION FEED MINERAL

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET CONT SHT

DYNAWEST PROJECTS LTD.


STEAM
378 t/h
917
VENT LOSS
34 t/h
919
WET STEAM
344 t/h
918
WATER
936
WT% t/h
WTR 99.0 7700
SOL 1.0 77
100.0 7777
OIL SAND FEED
197,196 t/sd
802
WT % t/h
BIT 12.1 994
WTR 5.4 443
SOL 82.5 6779
100.0 8216
EXTRACTION
BITUMEN RECOVERY
93.9%
REJECT
102
WT % t/h
BIT 1.3 3
WTR 13.7 34
SOL 82.5 210
100.0 247
TAILINGS
103
WT% t/h
BIT 0.4 58
WTR 54.8 8007
SOL 44.8 6549
100.0 14614
FROTH
101
WT % t/h
BIT 63.2 933
WTR 30.2 446
SOL 6.6 97
100.0 1476
WATER & DILUENT
204
WT% t/h
WTR 2.0 12
NAPH 98.0 606
100.0 618
STEAM
23 t/h
920
920
FROTH TREATMENT
BITUMEN RECOVERY
97.6%
TAILINGS
111
WT% t/h
BIT 4.4 22
WTR 76.1 393
SOL 17.0 88
NAPH 2.5 13
100.0 516
DILUTED BITUMEN
110
WT % t/h
BIT 57.0 911
WTR 5.5 88
SOL 0.6 9
NAPH 36.9 593
100.0 1601
DRY BITUMEN VOLUMETRIC
FLOW = 21,723 m³/sd

COMPOSITION	MATERIAL BALANCE - EXTRACTION								MATERIAL BALANCE - FROTH TREATMENT						
	OIL SAND	STEAM	WATER	TOTAL IN	FROTH	TAILINGS	REJECT	TOTAL OUT	FROTH	NAPHTHA	STEAM	TOTAL IN	DILUTED BITUMEN	TAILINGS	TOTAL OUT
BITUMEN	994			994	933	58	3	994	933			933	911	22	933
WATER	443	344	7700	8487	446	8007	34	8487	446	12	23	481	88	393	481
SOLID	6779		77	6856	97	6549	210	6856	97			97	9	88	97
NAPHTHA										606		606	593	13	606
TOTAL	8216	344	7777	16337	1476	14614	247	16337	1476	618	23	2117	1601	516	2117

t/h: TONNES PER STREAM HOUR

FIGURE 3.3.3
EXTRACTION AND FROTH TREATMENT MATERIAL BALANCE

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CONT. SHEET

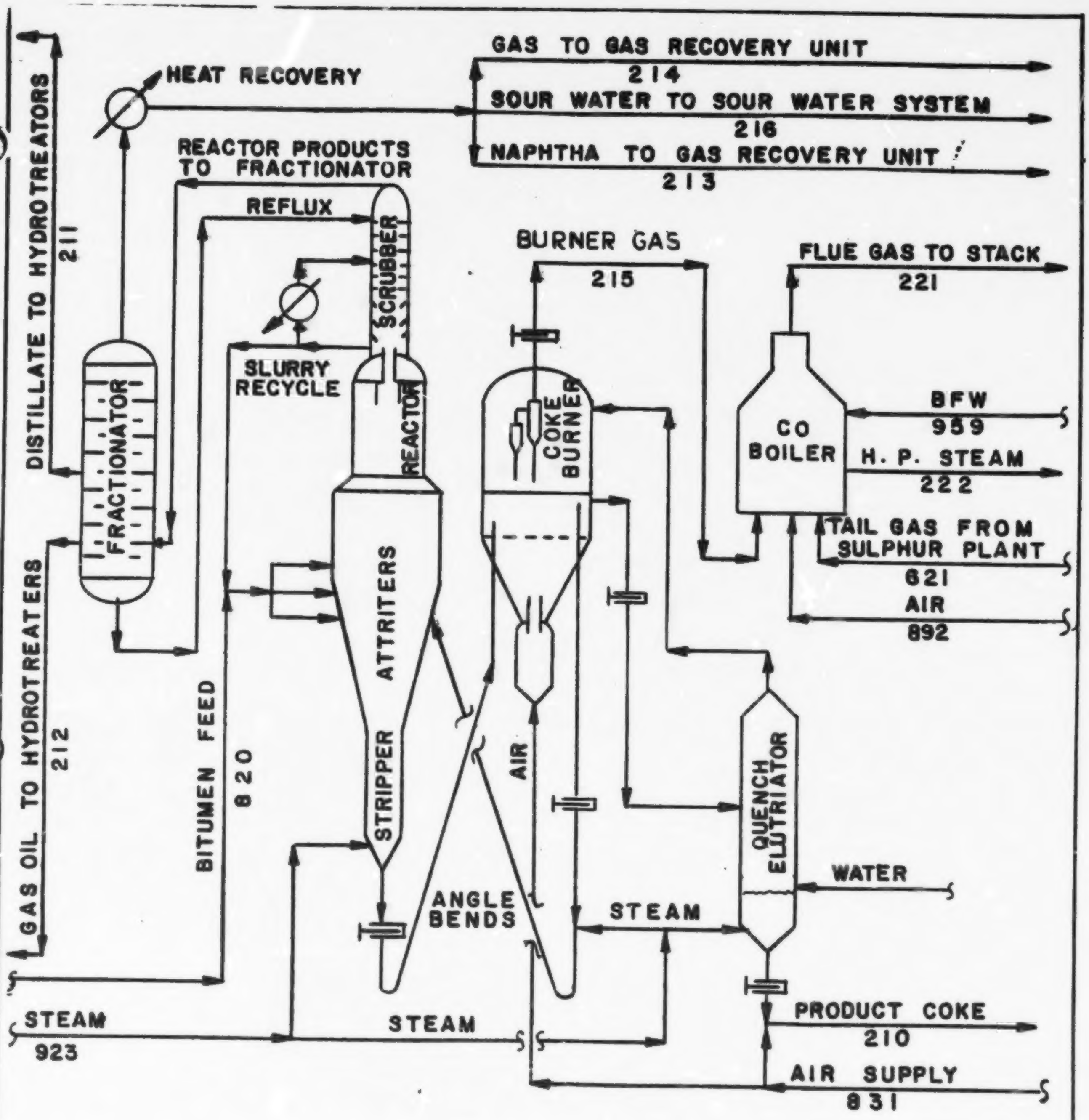
GAS TO GAS RECOVERY UNIT
214
HEAT RECOVERY
SOUR WATER TO SOUR WATER SYSTEM
216
NAPHTHA TO GAS RECOVERY UNIT
213
REACTOR PRODUCTS TO FRACTIONATOR
REFLUX
DISTILLATE TO HYDROTREATORS
211
SCRUBBER
BURNER GAS
215
FLUE GAS TO STACK
221
SLURRY RECYCLE
FRACTIONATOR
REACTOR
COKE BURNER
CO BOILER
BFW
959
H. P. STEAM
222
TAIL GAS FROM SULPHUR PLANT
621
AIR
892
ATTRITERS
GAS OIL TO HYDROTREATERS
212
BITUMEN FEED
820
STRIPPER
AIR
QUENCH ELUTRIATOR
WATER
ANGLE BENDS
STEAM
STEAM
923
STEAM
PRODUCT COKE
210
AIR SUPPLY
831

FIGURE 3.4.1
FLUID COKER PROCESS FLOW SCHEMATIC

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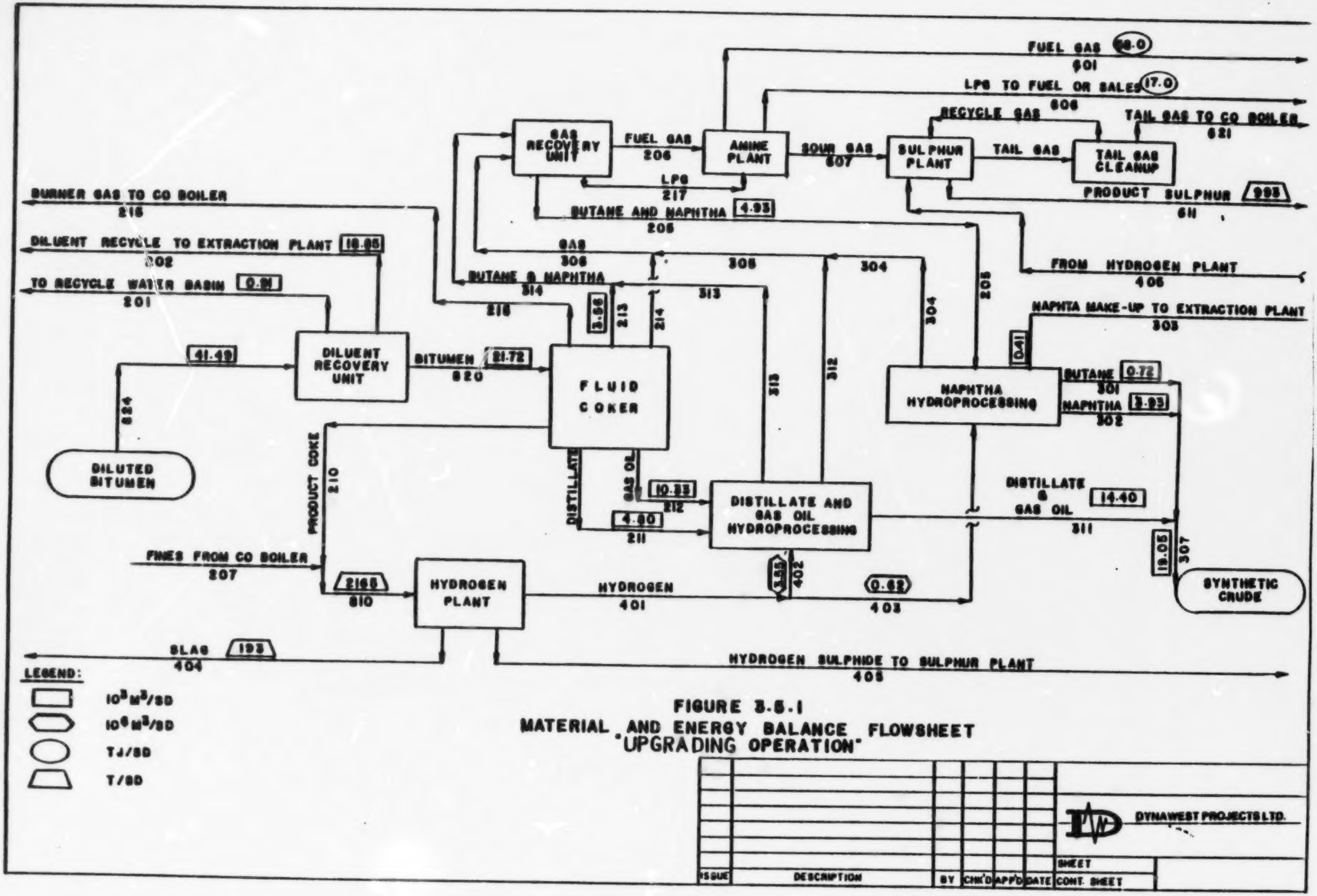
FUEL GAS
601
LPG TO FUEL OR SALES 17.0
606
RECYCLE GAS
TAIL GAS TO CO BOILER
621
GAS RECOVERY UNIT
FUEL GAS
206
AMINE PLANT
SOUR GAS
607
SULPHUR PLANT
TAIL GAS
TAIL GAS CLEANUP
LPG
217
PRODUCT SULPHUR 993
611
BURNER GAS TO CO BOILER
215
BUTANE AND NAPHTHA 4.93
205
DILUENT RECYCLE TO EXTRACTION PLANT 19.65
202
GAS
306
305
304
FROM HYDROGEN PLANT
406
TO RECYCLE WATER BASIN 0.91
201
BUTANE & NAPHTHA
314
313
216
3.56
213
214
205
304
NAPHTHA MAKE-UP TO EXTRACTION PLANT
303
0.41
41.49
824
DILUENT RECOVERY UNIT
BITUMEN 21.72
820
FLUID COKER
313
312
NAPHTHA HYDROPROCESSING
BUTANE 0.72
301
NAPHTHA 3.93
302
DILUTED BITUMEN
PRODUCT COKE
210
DISTILLATE
GAS OIL
10.33
212
4.80
211
DISTILLATE AND GAS OIL HYDROPROCESSING
DISTILLATE & GAS OIL 14.40
311
19.05
307
SYNTHETIC CRUDE
FINES FROM CO BOILER
207
2165
810
HYDROGEN PLANT
HYDROGEN
401
3.55
402
0.62
403
SLAG 193
404
HYDROGEN SULPHIDE TO SULPHUR PLANT
405
LEGEND:
$10^3 M^3/SD$
$10^6 M^3/SD$
TJ/SD
T/SD
DYNAWEST PROJECTS LTD.
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ISSUE
DESCRIPTION
BY
CHK'D
APP'D
DATE
CONT. SHEET

FIGURE 3.5.1
MATERIAL AND ENERGY BALANCE FLOWSHEET
"UPGRADING OPERATION"

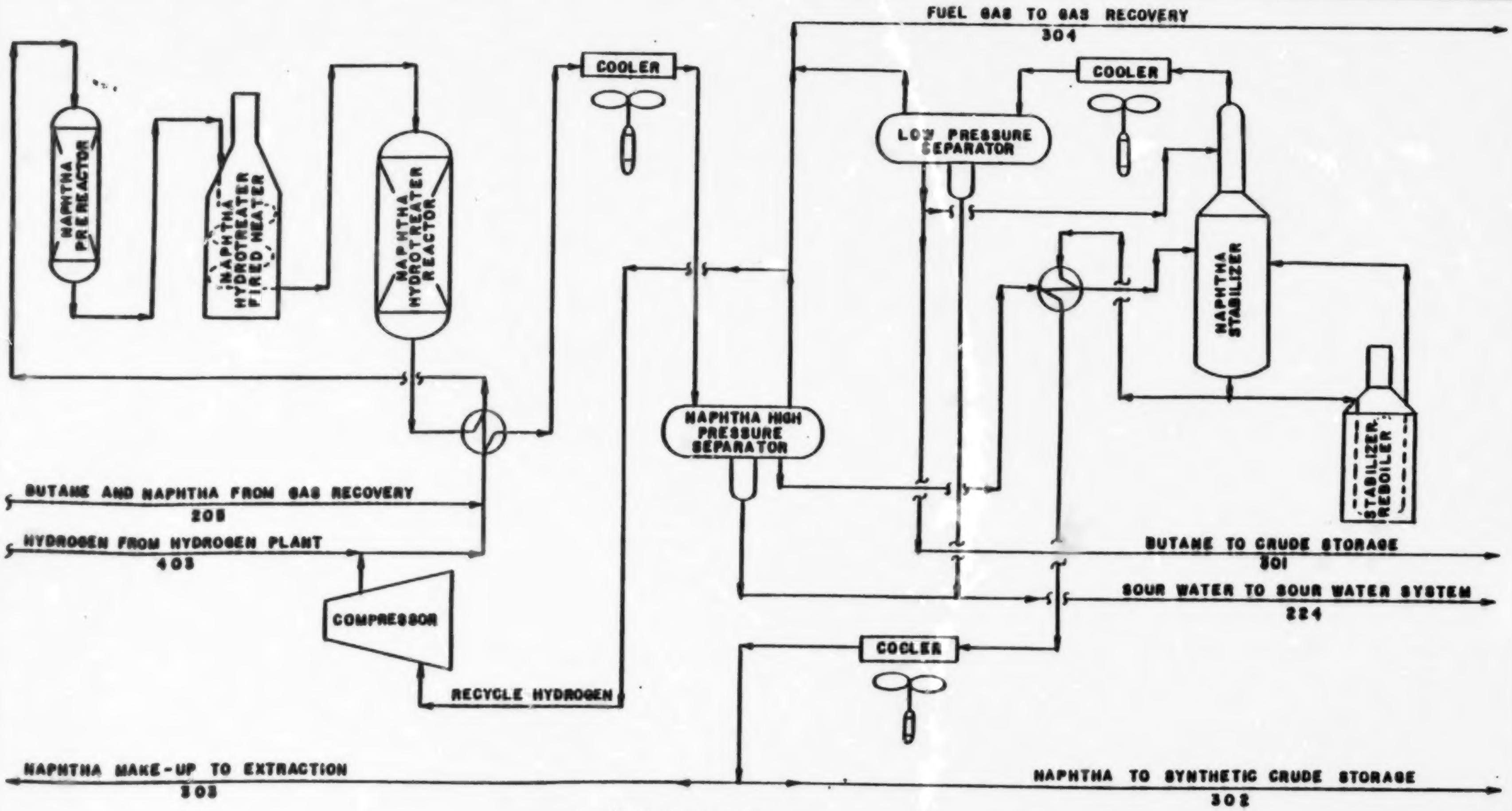
NAPHTHA PREREACTOR
NAPHTHA HYDROTREATER FIRED HEATER
NAPHTHA HYDROTREATER REACTOR
COOLER
NAPHTHA HIGH PRESSURE SEPARATOR
FUEL GAS TO GAS RECOVERY
304
LOW PRESSURE SEPARATOR
COOLER
NAPHTHA STABILIZER
STABILIZER REBOILER
BUTANE AND NAPHTHA FROM GAS RECOVERY
205
HYDROGEN FROM HYDROGEN PLANT
403
COMPRESSOR
RECYCLE HYDROGEN
BUTANE TO CRUDE STORAGE
301
SOUR WATER TO SOUR WATER SYSTEM
224
COOLER
NAPHTHA MAKE-UP TO EXTRACTION
303
NAPHTHA TO SYNTHETIC CRUDE STORAGE
302

FIGURE 3.5.2
NAPHTHA HYDROTREATER

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CONT. SHEET

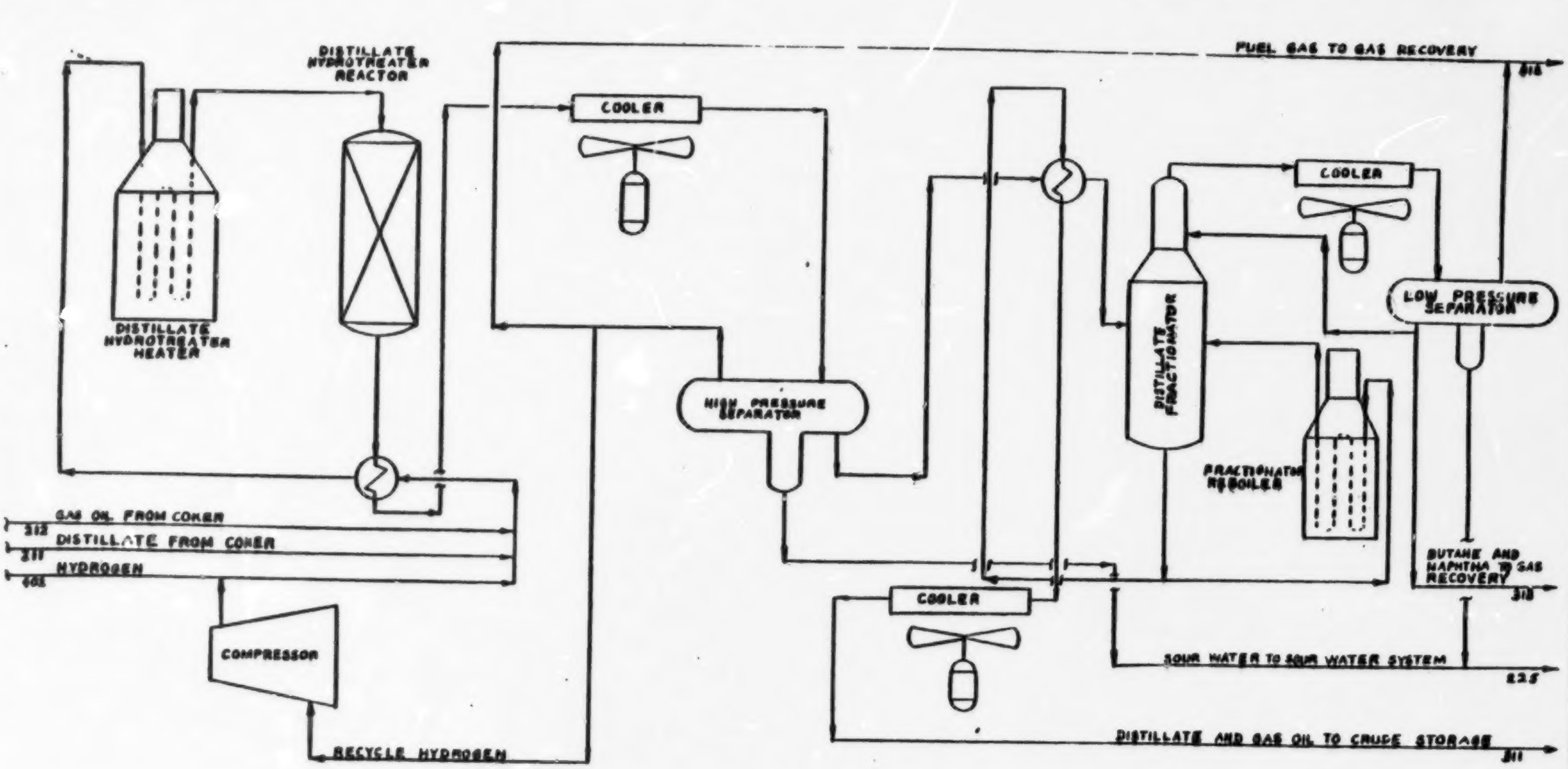
DISTILLATE HYDROTREATER REACTOR
DISTILLATE HYDROTREATER HEATER
COOLER
HIGH PRESSURE SEPARATOR
FUEL GAS TO GAS RECOVERY
316
COOLER
DISTILLATE FRACTIONATOR
FRACTIONATOR REBOILER
LOW PRESSURE SEPARATOR
BUTANE AND NAPHTHA TO GAS RECOVERY
310
GAS OIL FROM COKER
312
DISTILLATE FROM COKER
311
HYDROGEN
405
COMPRESSOR
RECYCLE HYDROGEN
COOLER
SOUR WATER TO SOUR WATER SYSTEM
225
DISTILLATE AND GAS OIL TO CRUDE STORAGE
311

FIGURE 3.5.8
DISTILLATE / GAS OIL HYDROTREATER

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

DYNAWEST PROJECTS LTD.

SHEET

CONT. SHEET

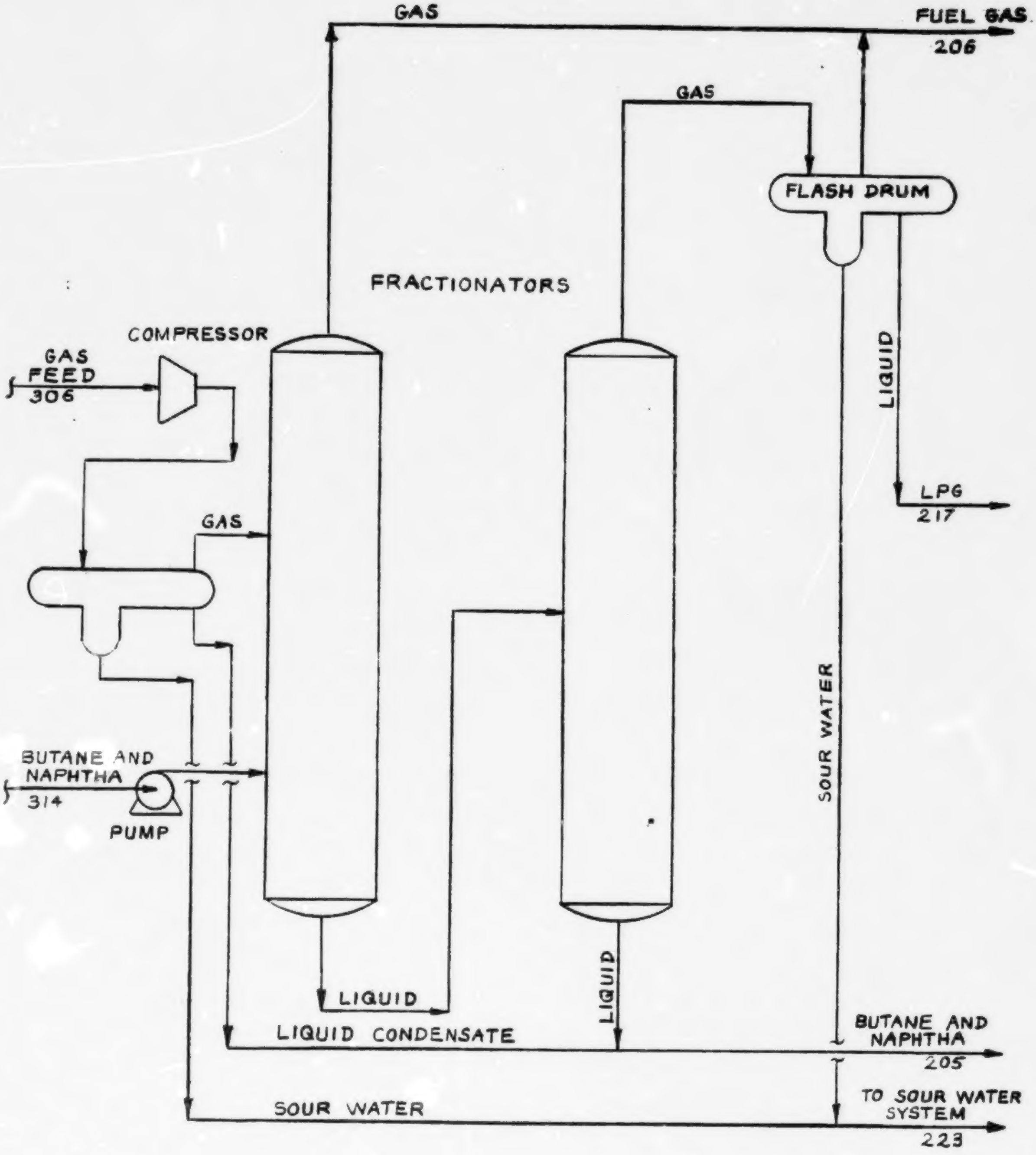
GAS
FUEL GAS
206
GAS
FLASH DRUM
FRACTIONATORS
LIQUID
COMPRESSOR
GAS FEED
306
LPG
217
GAS
SOUR WATER
BUTANE AND NAPHTHA
314
PUMP
LIQUID
LIQUID
LIQUID CONDENSATE
BUTANE AND NAPHTHA
205
SOUR WATER
TO SOUR WATER SYSTEM
223

FIGURE 3.5.4
GAS RECOVERY UNIT

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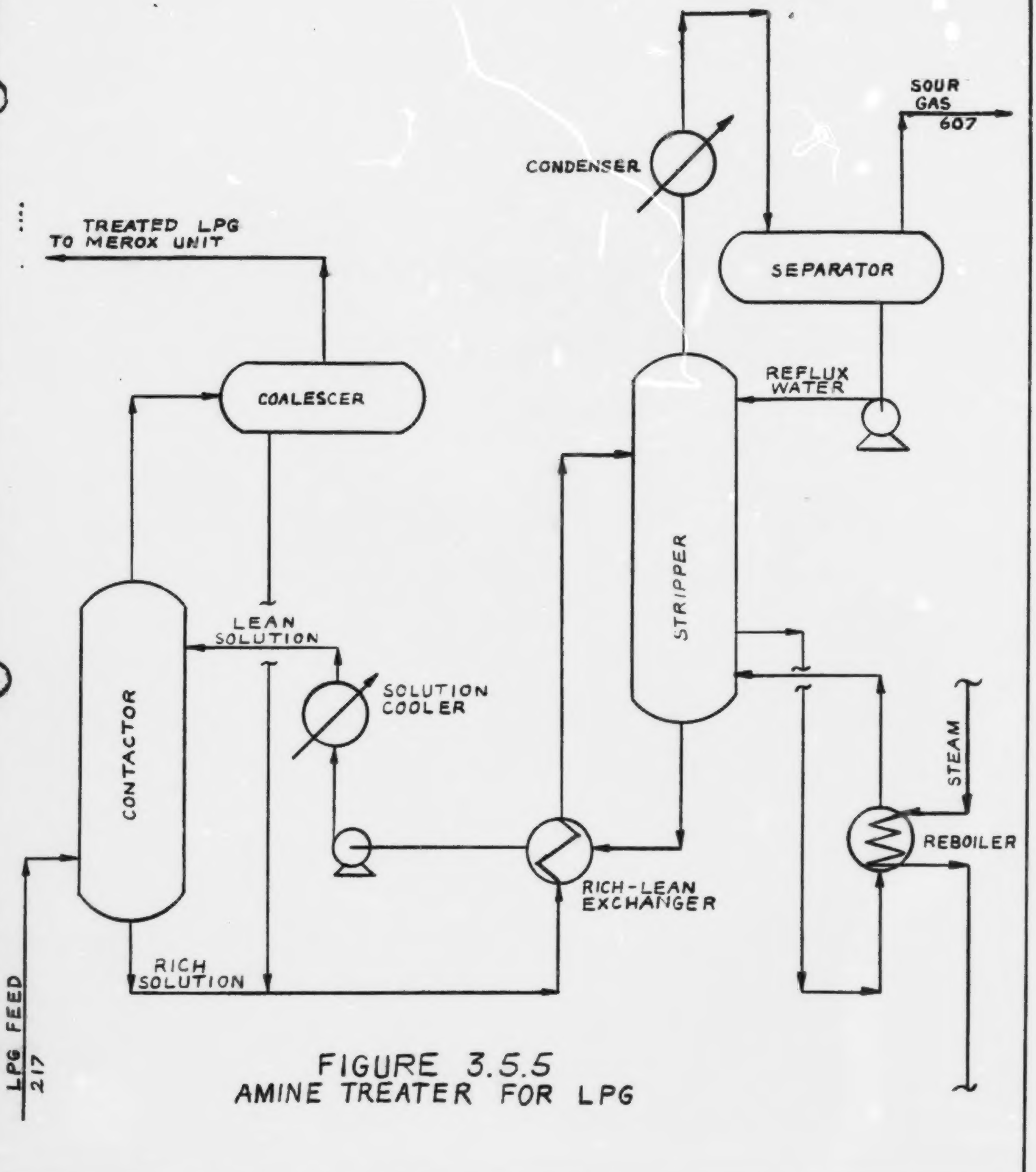
SOUR GAS
607
CONDENSER
TREATED LPG TO MEROX UNIT
SEPARATOR
COALESCER
REFLUX WATER
STRIPPER
LEAN SOLUTION
SOLUTION COOLER
CONTACTOR
STEAM
REBOILER
RICH-LEAN EXCHANGER
RICH SOLUTION
LPG FEED
217

FIGURE 3.5.5
AMINE TREATER FOR LPG

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CONT SHT

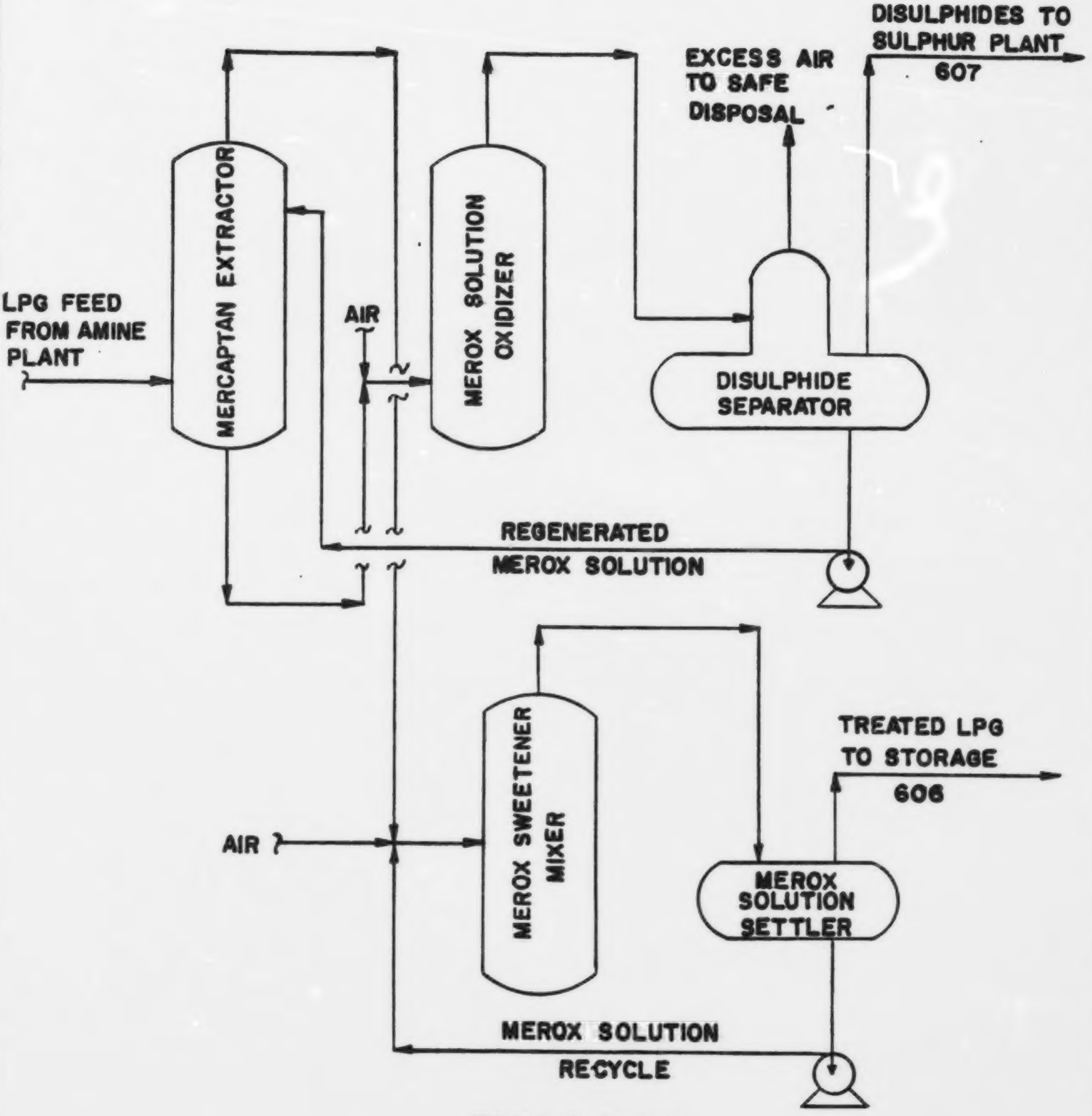
DISULPHIDES TO SULPHUR PLANT
607
EXCESS AIR TO SAFE DISPOSAL
MERCAPTAN EXTRACTOR
MEROX SOLUTION OXIDIZER
LPG FEED FROM AMINE PLANT
AIR
DISULPHIDE SEPARATOR
REGENERATED MEROX SOLUTION
MEROX SWEETENER MIXER
TREATED LPG TO STORAGE
606
AIR
MEROX SOLUTION SETTLER
MEROX SOLUTION RECYCLE

FIGURE 3.5.6
MEROX UNIT FOR LPG

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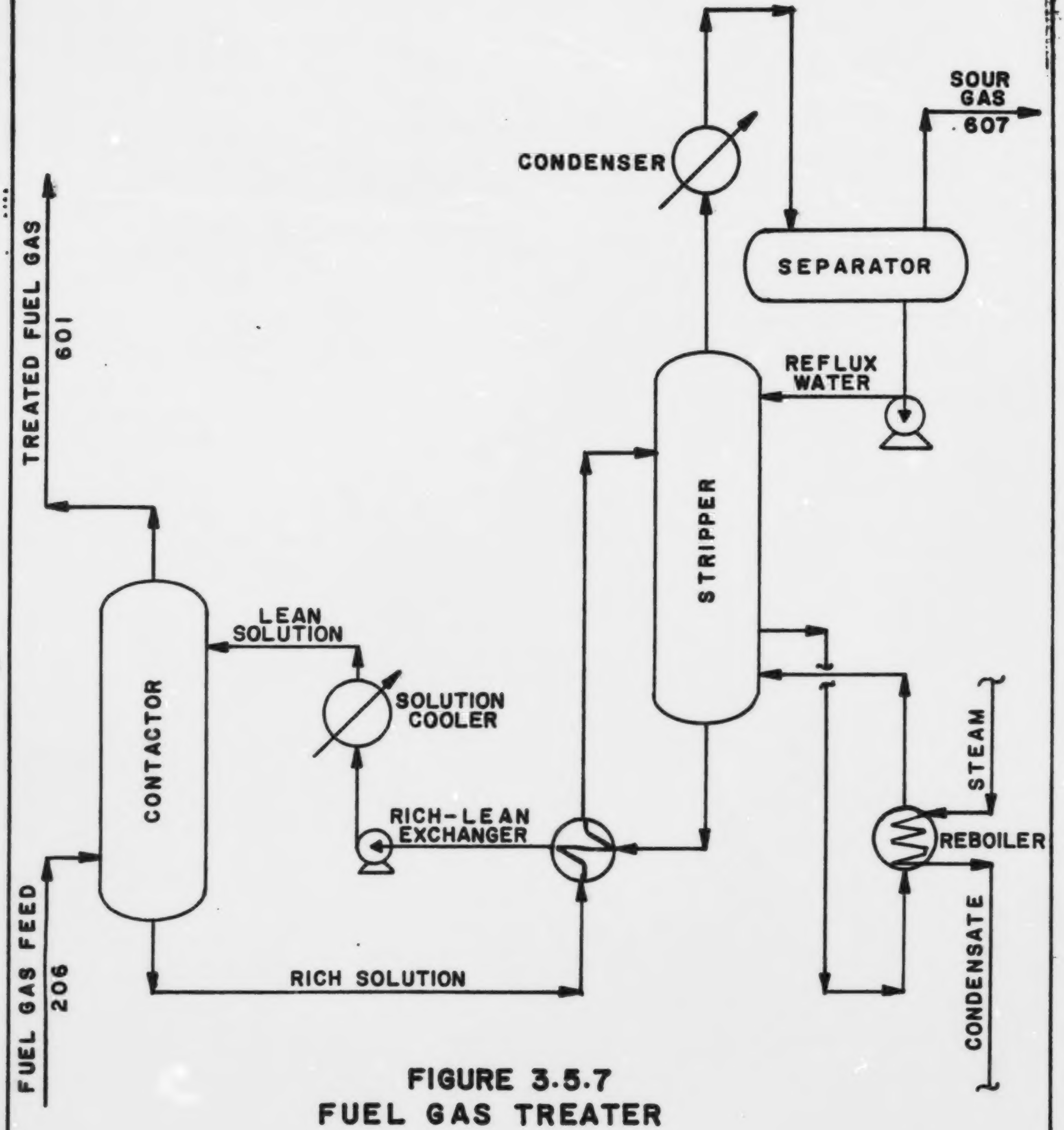
CONDENSER
SOUR GAS
607
SEPARATOR
TREATED FUEL GAS
601
REFLUX WATER
STRIPPER
LEAN SOLUTION
SOLUTION COOLER
CONTACTOR
STEAM
RICH-LEAN EXCHANGER
REBOILER
FUEL GAS FEED
206
RICH SOLUTION
CONDENSATE

FIGURE 3.5.7
FUEL GAS TREATER

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ISSUE DESCRIPTION BY CHK'D APP'D DATE CONT. SHT.

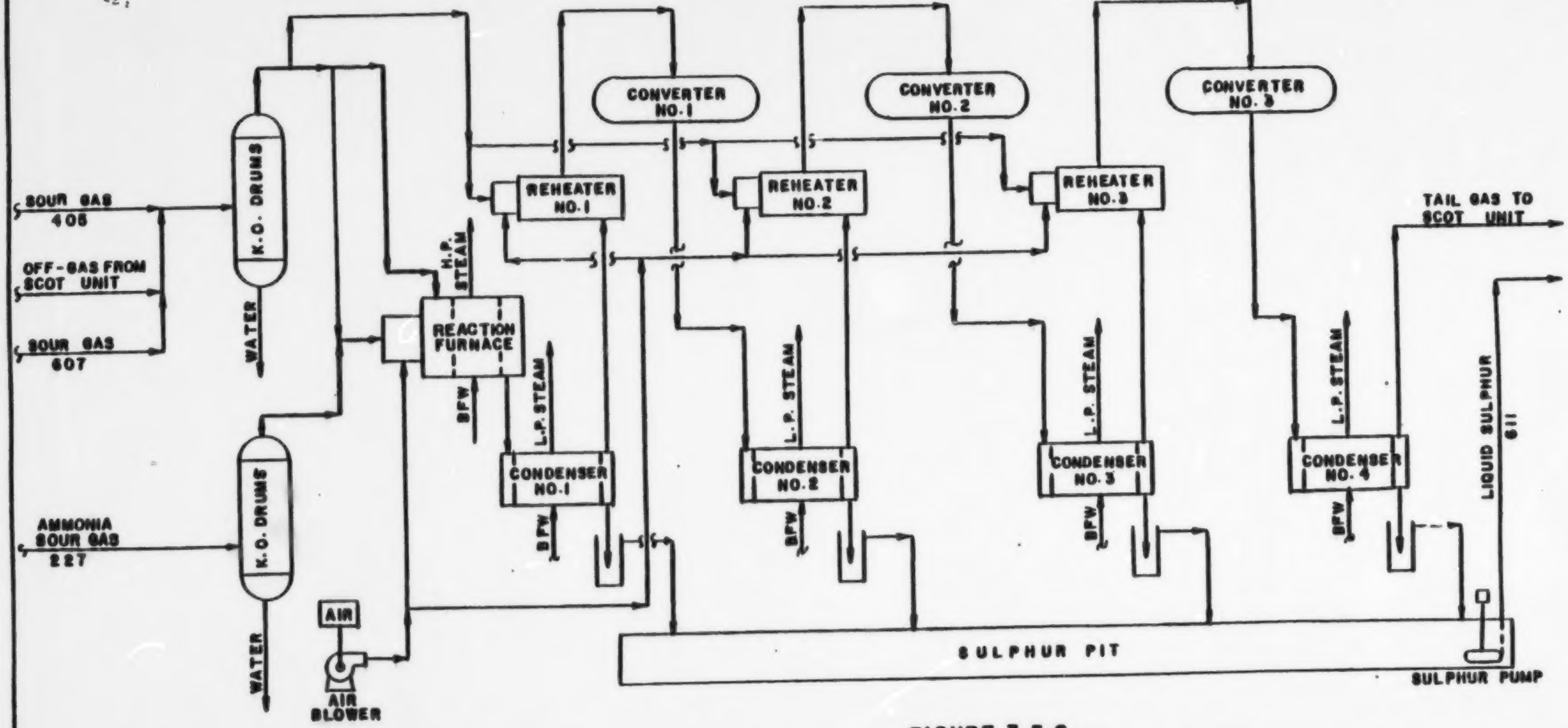
SOUR GAS
405
OFF-GAS FROM
SCOT UNIT
SOUR GAS
607
AMMONIA
SOUR GAS
227
K.O. DRUMS
WATER
K.O. DRUMS
WATER
AIR
AIR
BLOWER
REACTION
FURNACE
H.P.
STEAM
BFW
CONDENSER
NO.1
L.P. STEAM
BFW
REHEATER
NO.1
CONVERTER
NO.1
REHEATER
NO.2
CONDENSER
NO.2
L.P. STEAM
BFW
CONVERTER
NO.2
REHEATER
NO.3
CONDENSER
NO.3
L.P. STEAM
BFW
CONVERTER
NO.3
CONDENSER
NO.4
L.P. STEAM
BFW
TAIL GAS TO
SCOT UNIT
LIQUID SULPHUR
611
SULPHUR PIT
SULPHUR PUMP

FIGURE 3.5.6
THREE STAGE SULPHUR PLANT (CLAUS UNIT)

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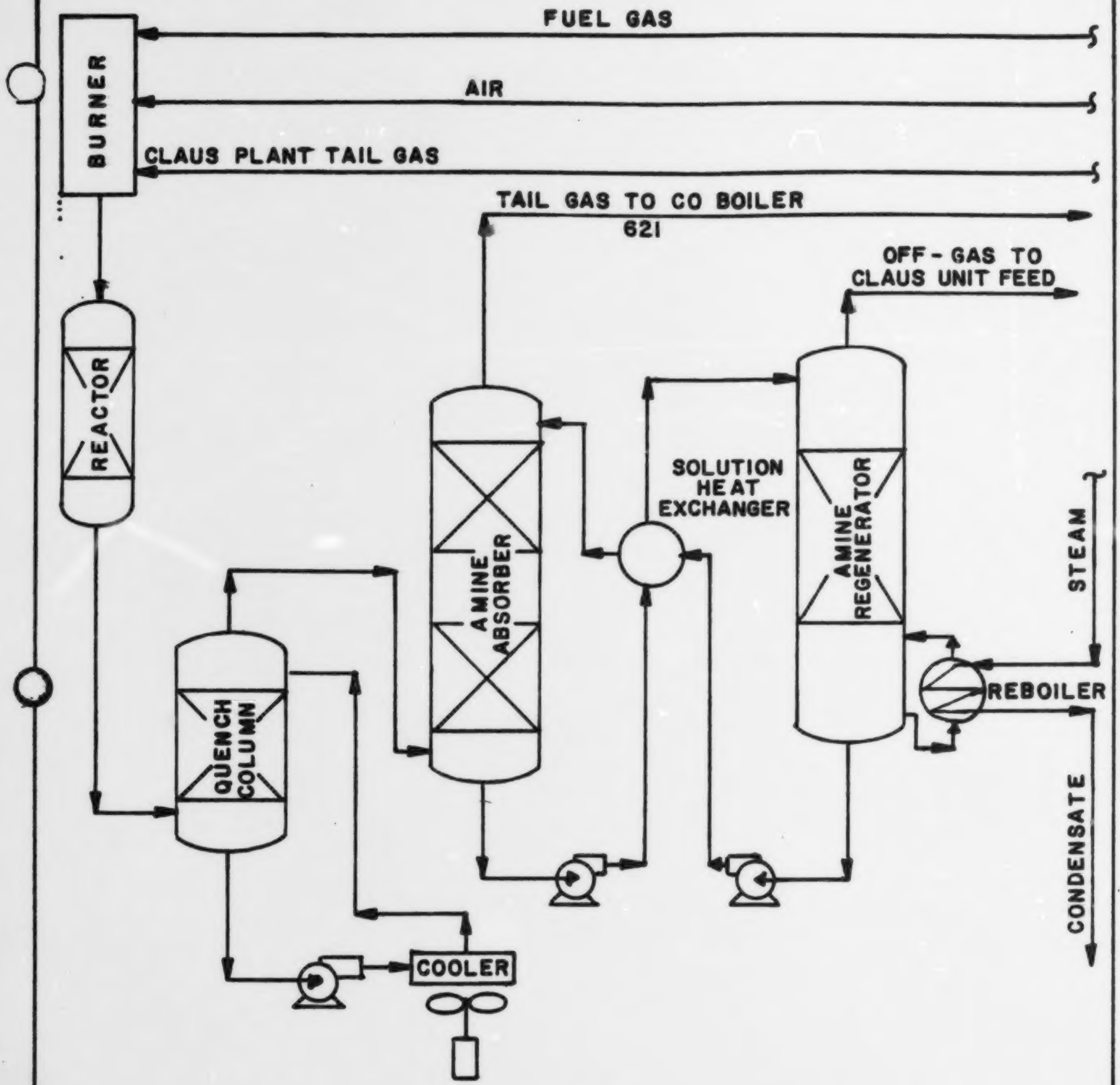
FUEL GAS
BURNER
AIR
CLAUS PLANT TAIL GAS
TAIL GAS TO CO BOILER
621
OFF - GAS TO
CLAUS UNIT FEED
REACTOR
SOLUTION
HEAT
EXCHANGER
AMINE ABSORBER
AMINE REGENERATOR
STEAM
REBOILER
QUENCH COLUMN
CONDENSATE
COOLER

FIGURE 3.5.9
SCOT TAIL GAS CLEANUP

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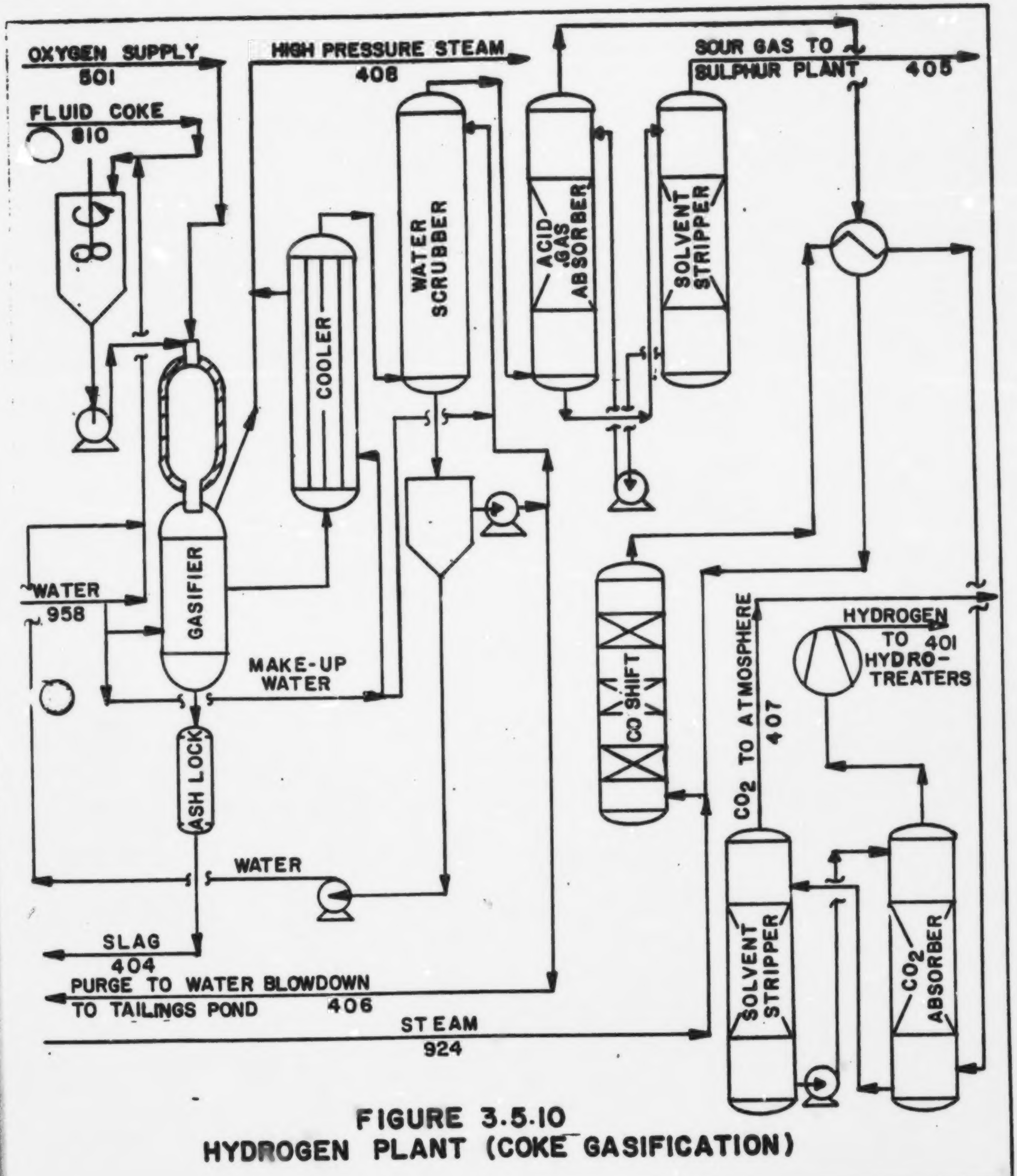
OXYGEN SUPPLY
501
FLUID COKE
810
HIGH PRESSURE STEAM
408
SOUR GAS TO
SULPHUR PLANT
405
WATER SCRUBBER
ACID GAS ABSORBER
SOLVENT STRIPPER
COOLER
GASIFIER
WATER
958
MAKE-UP
WATER
ASH LOCK
WATER
SLAG
404
PURGE TO WATER BLOWDOWN
TO TAILINGS POND
406
STEAM
924
CO SHIFT
CO_2 TO ATMOSPHERE
407
HYDROGEN
TO 401
HYDRO-
TREATERS
SOLVENT STRIPPER
CO_2 ABSORBER

FIGURE 3.5.10
HYDROGEN PLANT (COKE GASIFICATION)

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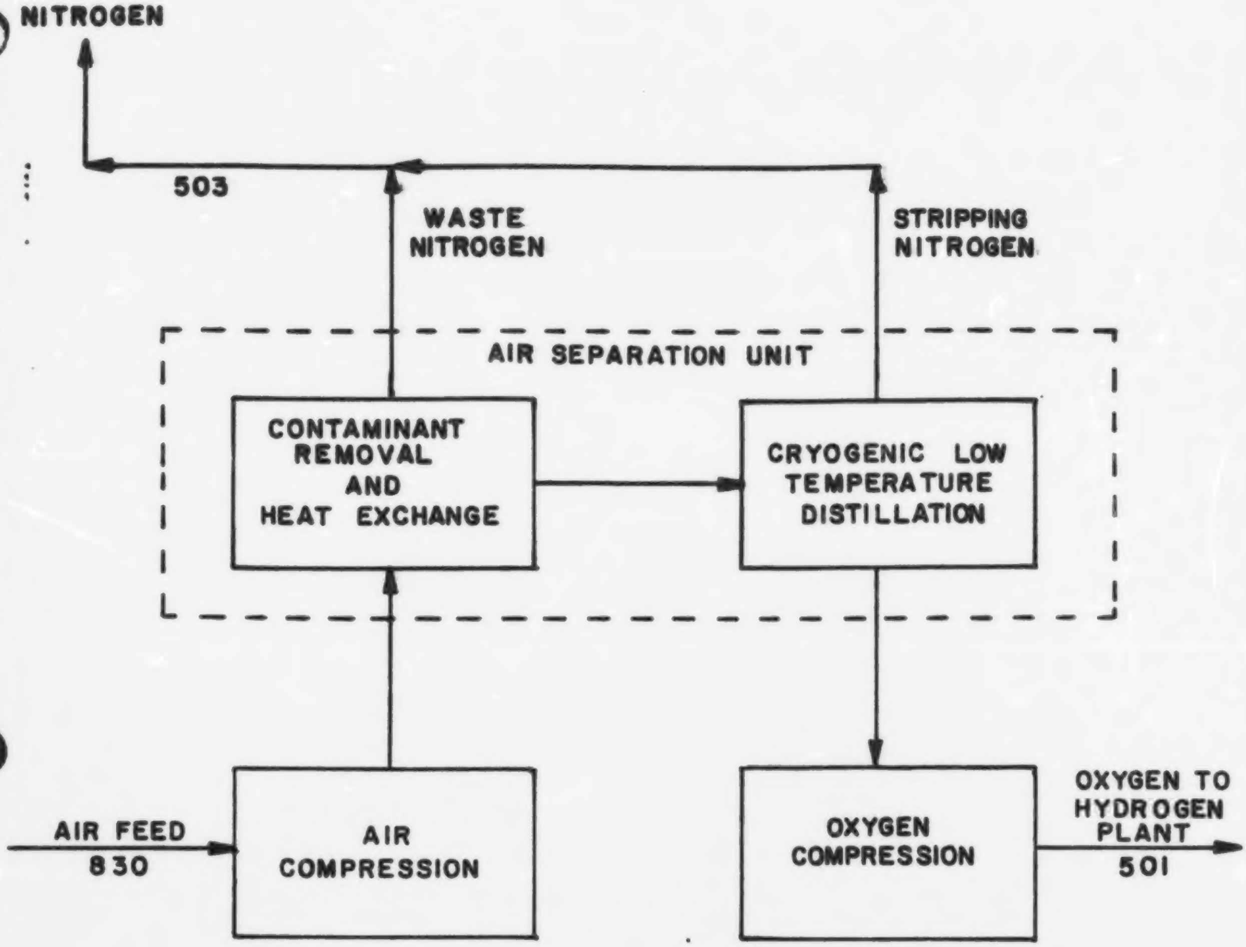
NITROGEN
503
WASTE NITROGEN
STRIPPING NITROGEN
AIR SEPARATION UNIT
CONTAMINANT REMOVAL AND HEAT EXCHANGE
CRYOGENIC LOW TEMPERATURE DISTILLATION
AIR FEED
830
AIR COMPRESSION
OXYGEN COMPRESSION
OXYGEN TO HYDROGEN PLANT
501

FIGURE 3.5.11
BLOCK FLOW DIAGRAM - OXYGEN PLANT

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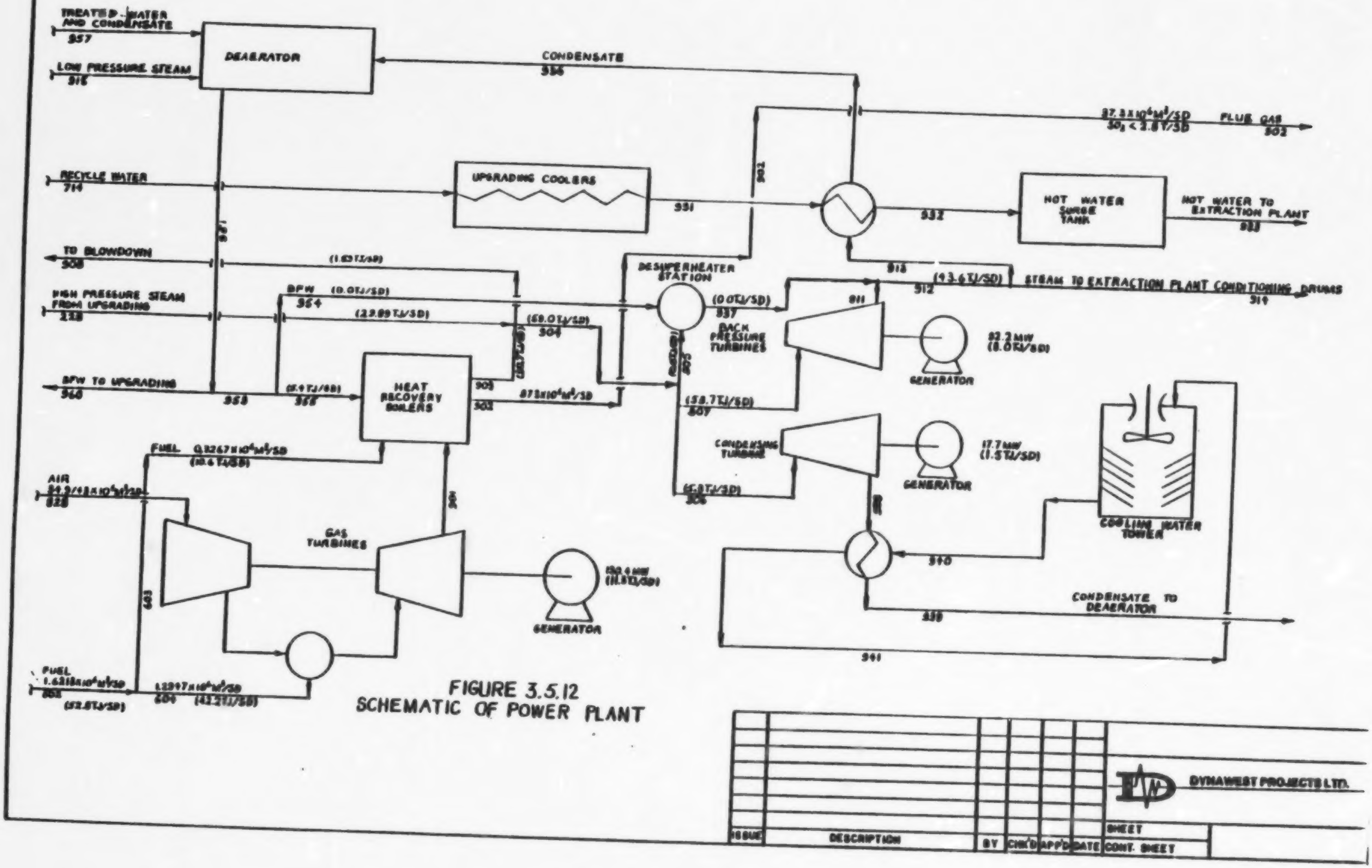
TREATED WATER AND CONDENSATE
357
LOW PRESSURE STEAM
316
DEAERATOR
CONDENSATE
336
RECYCLE WATER
714
UPGRADING COOLERS
331
TO BLOWDOWN
308
(1.8TJ/SD)
HIGH PRESSURE STEAM FROM UPGRADING
238
BFW (0.0TJ/SD)
354
(22.89TJ/SD)
(59.0TJ/SD)
304
BFW TO UPGRADING
360
353
(5.9TJ/SD)
355
HEAT RECOVERY BOILERS
303
302
873x10⁴M³/SD
FUEL 0.3267x10⁶M³/SD
(10.6TJ/SD)
AIR
323
GAS TURBINES
301
603
130.4MW
(11.8TJ/SD)
GENERATOR
FUEL
1.6218x10⁶M³/SD
602
(52.8TJ/SD)
1.2947x10⁶M³/SD
604
(42.2TJ/SD)
37.3x10⁶M³/SD
SO₂ < 2.8T/SD
FLUE GAS
302
HOT WATER SURGE TANK
332
HOT WATER TO EXTRACTION PLANT
333
315
(43.6TJ/SD)
312
STEAM TO EXTRACTION PLANT CONDITIONING DRUMS
314
DESUPERHEATER STATION
(0.0TJ/SD)
337
311
BACK PRESSURE TURBINES
82.2MW
(8.0TJ/SD)
GENERATOR
(58.7TJ/SD)
307
CONDENSING TURBINE
17.7MW
(1.5TJ/SD)
GENERATOR
(6.0TJ/SD)
306
COOLING WATER TOWER
340
CONDENSATE TO DEAERATOR
339
341
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APP'D
DATE
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FIGURE 3.5.12
SCHEMATIC OF POWER PLANT

4.0 THE PRODUCTION OF LIQUID FUELS FROM COAL USING THE HRI H-COAL PROCESS

4.1 INTRODUCTION

The H-Coal process for the production of liquid fuels is simply summarized on Figure 4.1.1. This shows a mine producing 64,600 t/d of raw coal which is crushed and dried to produce 44,500 t/d of dry coal feed to the liquefaction process. The liquefaction reaction produces 16,000 t/d of synthetic crude products together with fuel gas, sulphur, ammonia and phenols as byproducts. The major support units are two hydrogen plants, one utilitizing fractionation section vacuum tower bottoms, the other utilizing plant fuel gas as feeds. A sulphur plant treats all plant sour gases and a utilities plant provides liquefaction plant but not coal mine utility demands.

The operation is shown in considerably more detail on Figure 4.1.2. The flow rates and compositions for the streams shown on this figure are presented in Table 4.1. Detailed descriptions of each operation in the process are given in the sections that follow. Note that in the more detailed figures presented in subsequent sections all flowrates are referred back to Table 4.1.

4.2 THE MINING OPERATION

Sub-bituminous coal from the Alberta Plains Region is almost exclusively recovered by surface strip mining(21). The Picardville reserves are described as strip mineable and this is the case that will be described here. The main sources of data for this operation are publications(13, 15) which describe the Highvale mine on Wabumun Lake 70 km west of Edmonton. The Highvale mine will have an ultimate annual output of 14 million tonnes, i.e. about 75% of that required from the mine supplying the liquefaction plant.

In essence, the mining operation consists of the following stages:

1. Drainage.
2. Overburden removal.
3. Coal loading and hauling.
4. Partings removal.
5. Ash/slag disposal.
6. Reclamation.

Each stage is described below.

4.2.1 Drainage

Surface water from undisturbed areas adjacent to the mined area are diverted away from the pits by drainage ditches along the highwall. These ditches need progressive relocation as the mine advances. Active streams also need interception and diversion. Subsequent reinstatement of these streams would be included in the reclamation program.

In-pit water is channelled to sumps from which it can be pumped to ditches located on the highwall. This water is treated in retention ponds to remove particulates and floating contaminants such as oils and greases. Chemical treatment of discharge water is unlikely to be required although water quality is monitored to confirm this.

4.2.2 Overburden Removal

Overburden thickness varies typically from 20 m to 60 m. Prior to bulk overburden removal with draglines sufficient topsoil and subsoil are salvaged to provide 15 and 30 cm layers respectively over the spoil heaps. This material is generally recovered by scrapers and stockpiled for subsequent use in reclamation.

The remaining bulk of the overburden to a maximum depth of about 49 m is removed by draglines. Where overburden cover exceeds this depth it is reduced to 49 m by scrapers prior to stripping by dragline.

The precise operating method employed by the draglines is determined by a number of variables.

These include:

1. Overburden depth.
2. Dragline(s) capacity.
3. Pit width.
4. Stability of spoil heaps.

For a given mine the last three variables above tend to remain constant but the first, overburden depth, varies throughout the mine. This variation in overburden depth necessitates the corresponding use of different stripping configurations. Figures 4.2.1 and 4.2.2 show two typical configurations. Figure 4.2.1 illustrates single machine stripping without rehandle which is only feasible at depths down to about 20 m. Beyond this rehandle is necessary (i.e. a portion of the mined overburden has to be rehandled either by the same machine in a second position or by a second machine operating in tandem with the first. Figure 4.2.2 shows the position of the second machine which is situated on a levelled spoil heap previously cast by the first machine.

The majority of the overburden is removed without blasting, however, some areas of the mine may contain bands of sandstones and siltstones that require blasting. A detailed survey of the particular mine area is needed for a precise estimate of the amount of blasting required. In the Highvale mine it is estimated that about 14% of the overburden requires blasting with a low powder factor.

4.2.3 Coal Loading and Hauling

Thinner seams (1 m) are ripped and bulldozed into piles before being loaded with front-end loaders.

Coal in the thicker seams (3m) is drilled and blasted and the top 75 cm ripped with a dozer. It is then loaded with large electric shovels (typical capacity 11.5 m^3).

These loading operations are well proven in the coal industry. It should, however, be noted that since no downstream beneficiation is proposed (see Section 4.3 below) reasonable care would have to be taken to ensure that minimal overburden is loaded along with the coal.

Coal hauling can be by either conveyor or truck. A study for the Highvale mine showed that truck haulage is preferred. A typical fleet for this scale of mine would use 100 tonne bottom dump coal haulers.

4.2.4 Partings Removal

Partings are the material contained between the coal seams. At Highvale they are principally shale with some montmorillonite. there are two alternatives for partings removal. The first is to remove them from the pit area and leave them near the surface. In the case of the Highvale mine the high montmorillonite content would adversely affect the subsequent reclamation program by inhibiting root growth. The second is to leave the partings in the pit, however, this necessitates flatter spoil slopes and increases the amount of rehandle required by the draglines. Clearly a detailed analysis is required before a decision can be taken for a particular mine. At Highvale the partings are left in the pit.

Ripper dozers and scrapers are used to remove the partings.

4.2.5 Slag Disposal

Slag from the Texaco gasification unit will be disposed of in the mining pits. This can be achieved either by placing the slag on top of the partings at the bottom of the cuts or by placing it between the ridges in the spoil piles. In view of the sulphur content of the slag and its potential adverse effect on reclamation the former course is more likely.

4.2.6 Reclamation

After mining spoil piles and haul roads will be regraded to a maximum slope of 1 in 10. The subsoil and topsoil stockpiled during overburden removal will be spread over the regraded areas using scrapers. The recovered areas can then be harrowed, fertilized, re-seeded and reforested as required. At other mines in Alberta, alfafa, barley and rapeseed have been successfully grown on reclaimed land. Some areas have been returned to wildlife habitat by planting trees and shrubs.

It is anticipated that the land surface should be reseeded within three years of the initial mining disturbance. It is also anticipated that stockpiling of regrowth material, other than for haul roads would not be required for more than one year.

4.2.7 Equipment

Table 4.2.1 details the major items of equipment required for a typical mine producing the 19 million t/a coal required for this project.

4.3 COAL STORAGE AND PREPARATION

The operations details in this section are shown schematically in Figure 4.3.1.

The coal is supplied to the preparation unit in 100 tonne dump trucks. These vehicles load the coal into vibrating feeder drums which in turn deposit the coal onto a conveyor belt. This supplies a sampling station as well as a travelling boom stacker. At this point the coal is sent to either live or dead stoarge. The dead storage is a 30 day emergency coal storage pile which is covered with sod to reduce air permeation.

The live storage is an active storage pile with a 3 day capacity. Coal is removed from the active storage by an underground mobile reclaiming unit and fed to the mill supply conveyor which deposits the coal into mill supply hoppers. The coal, which has a moisture content of approximately 0.3 mass fraction, is crushed and dried to -20 mesh and 0.02 mass fraction water in a roller mill that is heated with hot combustion gases. The dried coal particles are pneumatically conveyed to cyclones and baghouses that separate the air and deposit the coal into storage bins. The storage bins are located at the end of the parallel mill trains and feed the slurry feed storage bins. These storage bins supply coal slurry mixing tanks where the coal is prepared for conversion.

The coal crushing and drying system processes large volumes of material. To maintain equipment of practical size, this part of the operation is divided into 21 parallel trains with an additional two trains provided as spares.

An additional unit that can be added to the coal supply system is a beneficiation system. This is a coal cleaning process that removes ash and/or sulphur. This is achieved by a water or chemical wash and can be tailored to a specific service depending on the properties of the coal being processed. Coal cleaning has certain associated costs and benefits. The benefits are the reduction of flue gas treatment and ash handling required while the costs are the increased capital expenditure and additional operating costs (including the increased load on the drying system).

To date, no cost benefit analysis of beneficiation in relation to coal conversion has been attempted(22). A detailed study would be required before this type of system could be considered for incorporation into this design.

4.4 COAL HYDROGENATION, PRODUCT SEPARATION AND UPGRADING

4.4.1 Reaction and Product Separation

The reaction section of the process comprises 27 parallel trains. This relatively large number is based on HRI's assumption that the first commercial H-Coal reactors will be approximately the same size as the proven H-Oil reactors from which they have been developed(23).

The main product separation and upgrading process comprises two parallel trains.

In Figure 4.4.1 which details these sections of the process only one stream is shown throughout for clarity.

The coal is fed from storage bins into slurry mixing tanks by screw feeders. It is mixed with solid free recycle slurry oil and the solids containing hydroclone overflow under an inert gas blanket to produce a 0.5 solid mass fraction slurry. Any vapours that are produced are condensed and combined with the slurry produced in the mixing tank and additional hydroclone overflow in the slurry holding tank. The slurry is maintained in uniform suspension by jet mixing and direct agitation. It is brought to reactor pressure (20675 kPa) by a centrifugal/reciprocating pump train. A portion of the preheated recycle hydrogen is injected into the process stream which is then heated (455°C) and introduced to the H-Coal reactor. The H-Coal reactor utilizes an ebullating bed of molybdenum cobalt catalyst that is supported by the upflow of coal-oil slurry, the balance of the preheated recycle hydrogen, as well as a vessel recycle stream that draws reagents from the top of the tower and reinjects them into the bottom of the tower. The reactor is shown schematically in Figure 4.4.2. The liquid and vapour effluent are carried overhead into the reactor effluent separator. The recovered vapour is partially condensed in hydrogen recycle and fractionator feed preheaters and then flows into the reactor vapour effluent separator; water is injected into the stream between the cooling stages to prevent the crystallization of ammonium sulphide salts. The cooled phase is split into vapour, slurry and hydrocarbon condensate components in the reactor effluent wash water separator. Part of the vapour overhead stream is compressed and combined with hydrogen produced in the cryogenic unit and recycled to the H-Coal reactor. Another portion of the vapour is combined with the vapours from the high pressure condensate wash water separator as feed to the cryogenic unit. The aqueous phase that contains the ammonium sulphide salt is processed in the water degassing drum.

The two stage condensation method of processing is used throughout the H-Coal process to remove ammonium sulphide salts and prevent the emulsification of the heavy hydrocarbon oils in the aqueous phase.

The slurry that is produced in the reactor and separated in the reactor effluent separator is fed to the high pressure slurry letdown drum where dissolved hydrogen is flashed off. The feed to the high pressure slurry letdown drum is cooled by direct contact with subcooled hydrocarbon condensate. This reduces the possibility of coke formation due to high temperature. The quench oil is supplied by a portion of the bottoms of high pressure condensate receiver and part of the bottoms of the fractionator lower side stripper which have been air cooled. The vapour produced is combined with the condensate from the reactor vapour effluent separator cooled and fed to the high pressure condensate receiver the vapour from which is injected with water, air cooled and combined with the hydrocarbon condensate from the reactor effluent wash water separator then fed to the high pressure condensate wash water separator. The vapour produced passes overhead to the cryogenic unit for further treatment while the aqueous product is fed to the water degassing drum.

The condensate produced by the high pressure slurry letdown drum is fed to the low pressure slurry letdown drum. The vapours produced in this flash vessel are combined with the remainder of the slurry produced in the high pressure condensate receiver and fed to the low pressure condensate receiver. The vapour collected from this flash vessel is injected with water and air cooled. It is combined with the hydrocarbon condensate produced in the high pressure condensate wash water separator and is then fed to the low pressure condensate wash water separator. The steam splits into three phases, an aqueous phase containing any ammonium sulphide salts that are sent to the water degassing drum, a vapour phase that goes to the sour gas clean up unit and a hydrocarbon condensate that is combined with the slurry from the low pressure condensate receiver and the hydrocarbon condensate from the low pressure wash water stripper as feed to the fractionator. The slurry from the low pressure slurry letdown drum contains all the solids in the H-Coal reactor product. The slurry is processed by the hydroclone system to reduce the solids concentration in the portion of the recycle slurry oil that controls the solids content in the reactor slurry. The hydroclone overflow is fed to the hydroclone overflow surge drum and then to the recycle slurry oil flash drum. The solids lean condensate is recycled to the slurry preparation tank. The hydroclone underflow is fed to the hydroclone underflow surge drum and then to the steam stripper. The bottoms from the steam stripper are fed to the vacuum tower where a distillate product is recycled to the slurry preparation tank. The vacuum tower bottoms are fed to the hydrogen plant (unit 400) for further processing. The steam stripper vapour product is combined with the vapour from the recycle slurry oil flash drum, cooled and then fed to the low pressure wash water stripper where the stream is split into three phases, the aqueous phase that is fed to the water degassifier, the vapour phase that is flared and the hydrocarbon condensate that makes up part of the fractionator feed.

The fractionator feed is made up of the combined hydrocarbon products from the steam stripper the low pressure condensate receiver and the low pressure condensate wash water separator. This feed is preheated in process heat exchangers and a fired preheater. The fractionator vapour overhead product is cooled and fed to the fractionator overhead drum that splits the flow into three phases, the aqueous phase that goes to the water degassing drum, the vapour product that goes to the cryogenic unit and the hydrocarbon product that goes to the naphtha stabilizer drum where it is combined with light condensate from unit 300. The vapour product passes to the cryogenic unit while the bottoms product passes on the the naphtha hydrotreater.

The fractionator has two product strippers. The lower produces a product that makes up a portion of the quench oil fed to the high pressure slurry letdown drum as well as supplying turbine oil for hydrotreating. The upper produces additional turbine oil that is fed to the hydrotreater. The fractionator bottom stream is a fuel oil, some of which is sent to hydrotreating, the remainder being recycled to the slurry preparation tank.

The water degassing drum is fed by numerous sources. The gases produced are vented to atmosphere while the sour water is sent to effluent control (unit 700) and the wet oil is treated for disposal.

4.4.2 Product Upgrading

The fractionator products contain unacceptable levels of nitrogen, sulphur and oxygen for subsequent refining and therefore require additional treatment.

The naphtha produced in the fractionation train is treated in the naphtha upgrading unit, shown schematically in Figure 4.4.2.1. It is combined with hydrogen from the hydrogen plant and then fed to a hydrotreating reactor preheater. The preheated stream is exposed to a fixed bed catalyst, typically a cobalt and/or molybdenum oxide. The products are then cooled and separated. The gaseous overhead product is rich in hydrogen and a portion is recycled to the hydrogen feed and the balance of this stream is used as fuel gas. The middle phase is fed to the light ends recovery unit. This fractionator produces a light ends product that is combined with the fuel gas product and then treated in a dedicated amine plant. The fractionator bottoms flow through the fractionator feed preheater and then to naphtha storage.

The turbine fuel and fuel oil are treated in similar equipment to that used for naphtha hydrotreating. Hydrotreating can virtually eliminate sulphur and nitrogen from the product; levels are determined by the severity of the reaction. Table 2.3.4 shows the quality of the hydrotreated products. Hydrogen requirements for product upgrading were based on data available in (17) and (18). It should be noted that much higher hydrotreating pressures (17 500 kPa) are required for H-Coal liquids than for oil sands liquids.

4.5 ANCILLARY PROCESS OPERATIONS

4.5.1 Light Ends Processing

The light ends processing unit is composed of two separate treatment systems each comprising two parallel trains (see Figure 4.5.1.1). The first processes vapour streams from the reactor effluent wash water separator and the high pressure condensate wash water separator. The second train treats the fractionator and stabilizer overhead as well as the vapour produced by the low pressure condensate wash water separator.

The first unit consists of a conventional amine plant, molecular sieve and a cryogenic unit. The amine process removes the hydrogen sulphide, carbon dioxide and ammonia that are contained in the hydrogen and light hydrocarbon gas. Hydrogen sulphide removal is required to meet environmental regulations when the fuel gas is burnt in the utilities plant and also to eliminate a build up of acid gas in the recycle hydrogen that is separated in the cryogenic unit. The carbon dioxide is removed to prevent solidification in the cryogenic unit.

The fuel gas is fed to the amine unit (Figure 4.5.1.2) where it is first treated with lean diethanolamine (DEA) solution. A sweet gas product goes overhead while rich DEA solution is fed to a reagent stripper via a rich-lean solution heat exchanger. The rich amine solution is stripped of the acid gas and is recycled to the contacting tower. The acid gas is fed to a Stretford plant (unit 600) for sulphur recovery.

The sweet gas is processed through molecular sieves to produce a clean, cool, dry gas that contains hydrogen and light hydrocarbon vapours. This stream is processed by a cryogenic unit to produce a purified hydrogen that feeds the H-Coal reaction as well as fuel gases for the rest of the process. The cryogenic unit consists of a multi-stage condensation in core type heat exchangers assisted by an external cascaded propane-ethylene refrigeration unit. This process utilizes a multiple stage flash at progressively lower pressures that lower the temperature of the system. The hydrogen that is recovered is recompressed and returned to the H-Coal reactor. The fuel gas produced is utilized in various areas of the plant.

The second treatment train consists of a diglycolamine (DGA) sweetening facility. The effluent vapour is compressed and any condensate is removed. The remaining gas is treated with DGA in the same manner as the gas in the first treatment train. The acid gas that is produced is sent to the Stretford plant (unit 600); the fuel gas is sent to the reforming unit the gas turbines and various process fired heaters.

4.5.2 Hydrogen Plant

The hydrogen required by the H-Coal reaction is supplied by the cryogenic plant and by this unit. Two systems of hydrogen production are used, a steam reforming process and a gasification system.

Steam reforming (Figure 4.5.2.1) processes the light hydrocarbon vapours (C_1 to C_5) that are separated in the DEA, DGA and cryogenic units. Any sulphur present in the feed is removed by a hydrodesulphurization catalyst, followed by zinc oxide adsorbent. The feed stream is preheated, combined with steam and fed to a tubular catalytic reforming furnace. Sweetened gas is used as a fuel for the furnace.

The product gas is rich in hydrogen and carbon monoxide. Additional hydrogen is generated by carbon monoxide conversion in carbon monoxide shift catalytic converters.

Carbon dioxide is removed in a hot potassium carbonate absorption unit. The overhead gas that is produced, mainly hydrogen, is compressed and combined with the hydrogen produced in the gasification unit for recycle to the H-Coal reactor.

The gasification unit (figure 4.5.2.2) uses the vacuum tower bottoms as feedstock for hydrogen generation. The vacuum tower bottoms are the residual portions of the H-Coal reaction, including high boiling point liquids, undissolved coal and ash. They have an initial boiling point in excess of 525oC and when partially oxidized produce a gas stream composed of H_2, CO, CO_2, H_2O, H_2S and small amounts of carbon.

Gasification of heavy low ash liquids is a fully commercial technology, however, high ash and solid or slurry feedstocks are just now undergoing commercialization. Texaco has gasified H-Coal vacuum tower bottoms at a pilot scale at its Montebello, California laboratories (65) A demonstration scale water-coal slurry fed Texaco gasifier was recently started-up in Alabama.

The vacuum tower bottoms are reacted with steam and oxygen in nine gasification reactors where partial oxidation occurs at a pressure of 5600 kPa and a temperature of approximately 1300^{o}C. This produces a two phase stream consisting of gaseous products, soot and molten ash. This high temperature stream is quenched with water and the ash is removed as a solid. The gas stream carries the soot into a scrubber where it is washed with water. The soot particles are picked up by the water and the gas is cooled by water evaporation. The clean gas is sent overhead to the carbon monoxide shift unit. The shift unit is similar to that used in the reforming section of unit 400. A cobalt molybdate catalyst that is resistant to sulphur poisoning is used.

The product gas is treated in a Selexol acid gas removal process (Figure 4.5.2.3). The hydrogen sulphide and carbon dioxide concentrations are sufficiently low that the first stage of the Selexol process could probably be removed at the expense of a much larger residual gas flow to the sulphur plant.

4.5.3 Oxygen Plant

Conventional technology is used to supply the gasification unit with oxygen. The cold box system that is used also supplies the nitrogen for regenerating the acid gas absorbent as well as the inert gas blanket used in the slurry preparation unit.

The oxygen unit (Figure 4.5.3) filters atmospheric air and compresses it to approximately 625 kPa. The air is cooled by heat exchange with product oxygen and nitrogen. Nitrogen is periodically flushed through this portion of the system to remove any impurities. A cryogenic distillation is used to separate nitrogen and oxygen. The purified oxygen is then compressed to approximately 7600 kPa and fed to the gasification system (unit 400).

4.5.4 Emissions Control System

The emissions control system (figure 4.5.4.1) reduces the amount of sulphur that is released to the atmosphere. It consists of the following components:

a Claus sulphur recovery unit (figure 4.5.4.2)
a Beavon-Stretford unit (figure 4.5.4.3)
and a sulphur flaking system.

The Claus plant (figure 4.5.4.2) recovers pure sulphur from the acid gas produced in the ammonia recovery process (unit 700). This stream is fed to a split flow Claus plant where a portion of the hydrogen sulphide is converted to sulphur dioxide in a waste heat boiler. The sulphur dioxide is then exposed to an activated alumina or bauxite catalyst and the remaining hydrogen sulphide to produce elemental sulphur, which is then condensed. The molten sulphur is cooled, flaked and stored for disposal.

The Claus tail gas is fed to a Beavon unit (figure 4.5.4.3) where the sulphur is catalytically hydrogenated to hydrogen sulphide. This stream is then mixed with the acid gases recovered in the amine and Selexol plants and fed to a Stretford absorber. The Stretford process exposes the hydrogen sulphide to sodium carbonate anthraquinone disulphonic acid and vanadium salts that absorb the impurities. The hydrogen sulphide is then oxidized to elemental sulphur in an aeration system and removed from the absorbing solution by a centrifuge. The sulphur is melted, then flaked and stored for disposal. The sweetened effluent gas is released to the atmosphere.

The Claus-Beavon-Stretford combination is used because of the variation in hydrogen sulphide concentration in the acid gases. The acid gas from the ammonia recovery unit has a much higher sulphur concentration than that found in the gases from the amine and Selexol plants. The Claus plant is suited to the higher sulphur concentrations and is therefore used to treat the acid gas from the ammonia plant while the Beavon-Stretford plant is used to treat the Claus tail gas and the remaining acid gases from other plant units.

4.5.5 Effluent Control Systems

The sour water produced in the coal liquefaction process is treated to remove oil, phenol, sulphur and ammonia (figure 4.5.5.1).

The aqueous stream is fed to a sour water stripping/ammonia recovery unit. Hydrogen sulphide and anhydrous ammonia are recovered from the sour water stream. These contaminants are separated from each other by the Phosam-W process which is currently used in commercial operations. The stripped water is processed in a phenol recovery unit. Proprietary steel industry technology that is commercially proven is used in this unit. The treated water is recycled for plant consumption.

The ammonia and phenols recovered in this process are sent to storage for subsequent sale.

4.6 UTILITIES

The utilities plant supplies the coal liquefaction facility with the energy required for plant operation. Power required at the mine site will be supplied by either the provincial electric grid or by natural gas fired generation facilties. Production and consumption of major utilities is summarized in Table 4.6.

4.6.1 Raw Water Storage and Filtration

The coal liquefaction process consumes approximately 75,000 m^3/d of raw water. This is the maximum requirement that occurs during the summer months when the cooling system evaporation losses are greatest.

The water is supplied through a pipeline from either the North Saskatchewan or Pembina Rivers. The water is stored in a reservior with a one day residence time. This provides a water supply for firefighting and reduces the possibility of interruption of supply.

The water is cleansed in granular filter beds that remove small amounts of sediment prior to being added as boiler feedwater makeup. Any backwash is sent to the cooling tower systems.

4.6.2 Boiler Feedwater

The water that is used as boiler feed water is processed in a treatment scheme that provides the high purity required for the production of 10,500 kPa steam. The water is treated initially in a micron filter that removes traces of suspended solids it is then processed by a reverse osmosis unit that removes about 95% of the dissolved solids. The remaining contaminants are removed in a conventional mixed bed ion exchange system. The demineralized water is then sent to storage and deaeration.

4.6.3 Cooling Tower and Cooling Water System

Two cooling water systems are used in this plant, a clean water system and an oily water system. The clean water system serves the utility and offsites area. Oil free raw water and some of the treater aqueous effluents are the only makeup streams that are acceptable.

The oily cooling water system is used in the balance of the plant. This system is oil contaminated and receives makeup from the utilities cooling system, filter backwash, reverse osmosis concentrate and deoiled oily water wastes among others. These streams are treated with a cold lime-soda process to reduce scaling.

Sulphuric acid is used for pH control and several proprietary chemicals are used for system maintenance.

4.6.4 Steam and Power Generation

Electricity is supplied to the plant by a combined cycle power plant. The plant is composed of a gas turbine-electric generator, a heat recovery steam generator and a steam turbine-electric generator.

The gas turbines are fueled with gas recovered from the process and supplemental natural gas. All the power produced is consumed within the plant, eliminating the need for external power consumption. A spare power generation train is available in case of equipment failure.

The heat recovery steam generator recovers the heat contained in the gas turbine exhaust by producing steam at 10,500 kPa and 450 kPa. this reduces the exhaust temperature to 175°C as it is vented to the atmosphere. The 10,500 kPa steam generated in the heat recovery steam generator is combined with steam produced in the process area and is superheated to 455°C.

A portion of this steam is consumed in the power generation steam turbines. The balance is combined with 4,225 kPa steam for process consumption. Excess 4,225 kPa steam is injected interstage in the steam turbine. Excess low pressure steam that is not utilized in the main process is used for power generation in low pressure turbogenerators.

4.6.5 Steam and Condensate System

Steam is supplied to the plant at four pressures, 10,500, 4,225, 1,135 and 450 kPa. The power generation units produce sufficient steam for plant requirements so conventional steam boilers are not required.

The makeup water is treated with protective chemicals that reduce the possibility of solids deposition as well as pH and dissolved oxygen control.

4.6.6 Fuel Gas System

The fuel gas produced in the cryogenic unit supplies two systems. A portion is fed to the process fired heaters at 345 kPa, while the balance is used as feedstock for the reformer at 2,720 kPa. The balance of the fuel gas generated in the plant is fed to the oxygen and power plants.

4.7 OFFSITE FACILITIES

4.7.1 Tankage and Shipping

Storage for two weeks production is required. Naphtha is stored in floating roof tanks because of the high vapour pressure, while fixed roof tanks are used for lower volatility liquids. Refrigerated spheres are used for ammonia storage.

Product tankage requirements are summarized below:

Naphtha	127,000 m^3
Turbine Fuel	88,000 m^3
Distillate Boiler Fuel	53,000 m^3
Ammonia	5,500 tons
Phenols	800 tons
Sulphur	6,000 tons (stockpile)

Vapour recovery systems are used on the storage vessels to minimize losses.

Storage capacity of 125,000 m^3 is also provided for the sour water produced in a one week period to isolate the process from a sour water system shutdown.

Additional storage of 185,000 m^3 is provided for intermediate products and off specification material. This volume was chosen to maintain consistency with the oil sands portion of this report.

The waste water produced in the reverse osmosis unit is sent to the cooling water system while the demineralizer waste is sent to disposal.

4.7.2 Potable Water

The utility system supplies potable water throughout the plant. The water is filtered and chlorinated to the levels set by Alberta Environment for municipal water supplies.

4.7.3 Waste Water Treatment

The aqueous product from the sour water stripper may contain traces of hydrocarbon. These are removed and the remaining portion is processed by the phenol recovery unit.

Any oily water from the storm basin forebay is treated for oil extraction and then filtered in a deep bed unit. The clean water is recycled to cooling system makeup water stream.

Storm water that is collected in areas where water contamination could occur is diverted with the utility waste water into the main basin. Water from locations where contamination could not occur is sent directly to the holding basin.

4.7.4 Flare and Blowdowns

All units in the facility are provided with appropriately sized pressure relieving devices and blowdown headers which are connected to the flare system. The flare system is provided with an automatic ignition system and knock-out drums.

4.8 ENVIRONMENTAL IMPACT

4.8.1 General

The proposed coal liquefaction facilities are located approximately 40 km north of Edmonton. The proximity to populated areas necessitates strict environmental control measures.

4.8.2 Surface Disturbances

The two main sources of surface disturbance are:

- the strip mine;
- the plant site.

The mining procedure is described in Section 4.2. The strip mine will cover approximately 75 km^2. The total area will not be exposed for the duration of the operation but in sections as the project progresses. The topsoil and overburden that cover the coal disposit are removed. The topsoil is stored for future revegetation while the overburden is cast directly back into the mine. As these areas are regraded they are covered with topsoil and revegetated. The land reclamation trails the excavation by approximately two years. It is believed that the ground cover will be re-established within five years.

The plant site will cover approximately 25 km^2. This will be used throughout the project and will be reclaimed at the end of the project.

4.8.3 Water Management

The disposal of surface and underground water is a major environmental concern. The coal seams that are removed tend to duct water into the mined out area. This water and any other underground or surface water is diverted into ditches and is processed through settling ponds that remove any suspended solids. The clarified water is disposed of in natural waterways.

The coal liquefaction process consumes approximately 75,000 m^3/d of process water. This water must be replaced from either the Pembina or North Saskatchewan River. The Pembina is the closer of the two but it may not have enough capacity to satisfy the plant requirements.

Approximately 85% of the makeup water requirement is needed because of evaporation loss from cooling towers. This loss could be reduced by implementing alternative cooling systems.

A detailed study regarding the cooling requirements and the water source suitability is needed before a cooling strategy can be developed. Such a study would need to consider heat optimization, fluctuating water flow rates as well as other relevant criteria.

The process water becomes contaminated with various pollutants, particularly phenols, hydrogen sulphide, carbon dioxide and ammonia. The aqueous stream is treated in a sour water stripping system that recovers ammonia and hydrogen sulphide as separate streams. The phenols can be removed by a number of proprietary recovery techniques. Free phase hydrocarbons are separated by a simple gravity decantation process.

These water treatment processes remove pollutants to a level that satisfies the standards set by the Alberta Government.

4.8.4 Atmospheric Emissions

The harmful emissions produced by the coal liquefaction process are mainly sulphur based.

The coal is fed to a reactor-separator-fractionator train. The separator and fractionator both produce fuel gases that are treated in conventional amine plants. The acid gases that are produced by the amine plants are combined with the tail gas from the hydrogen facility and are treated in Stretford units which convert the sulphur compound to a solid sulphur product. The liquid product from the reactor-separator-fractionator unit is processed in hydrotreaters where a hydrogen sulphide product is formed. This and the hydrogen sulphide produced in the water treatment facility are treated in a Claus plant. The tail gases from the Claus and Stretford plants are combined and released to atmosphere through a stack.

The Claus plant processes approximately 75% of the sulphur that is removed while the Stretford plant treats the remaining 25%. This reduces the sulphur concentration to approximately 200 x 10^{-6} volume fraction of sulphur dioxide. Trace amounts of hydrogen sulphide are released but the concentration is sufficiently low (40 x 10^{-6} volume fraction) to be vented to the atmosphere.

The dust and noise produced by the strip mining operation are another area of concern. The proposed solution to this problem is the development of a contoured sight and sound barrier. This, in conjunction with a dust suppressant, has been proven, in similar operations, to reduce this problem to a minimum.

4.8.5 Service Network

The proposed site for this project is located 40 km north of Edmonton. The access and power to the area are already in place so no major additional development of the infrastructure should be necessary.

The most damaging component of the service network will be the installation of a pipeline that will supply the makeup process water. At present the water source has not been determined. The Pembina or North Saskatchewan River will probably be the source but a more detailed study is required to determine which would be most beneficial. The majority of the personnel required by the plants are expected to commute from Edmonton, which will ameliorate the social impact of the coal liquefaction facility.

4.8.6 Conclusion

The coal liquefaction facility will affect a large area of land. The proximity to urban areas necessitate strict air and water pollution control. State of the art technology is capable of reducing the contaminants to levels well below the standards that are set by the Alberta Government.

With responsible land reclamation and the application of good pollution control technology, the environmental impact of this coal liquefaction facility should be minimal.

NUMBER / STREAM / COMPONENT	COMPONENT PROPERTIES MOL WT	SG	°API	101 RAW COAL FROM STORAGE TO CRUSHING AND DRYING	102 CRUSHED AND DRIED COAL TO SLURRY	201 COAL SLURRY TO H-COAL REACTOR	202 LIQUEFIED COAL FROM H-COAL REACTOR	203 SOUR GAS FROM [illegible] TO AMINE PLANT	204 SOUR FUEL GAS FROM EFFLUENT SEPARATOR	205 [illegible] TO FRACTIONATION	206 SLURRY TO [illegible]	207 VAPOUR TO FRACTIONATION	208 SOUR WATER FROM EFFLUENT SEPARATOR	209 OIL RECYCLE TO COAL SLURRYING	210 [illegible] RECYCLE [illegible] EFFLUENT SEPARATOR	211 HYDROGEN RECYCLE TO H-COAL REACTOR	212 SOUR GAS [illegible]	213 [illegible] FUEL GAS TO AMMONIA PLANT	214 NAPHTHA TO NAPHTHA HYDROPROCESSING
				m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)
PURE COAL				1905	1905														
WATER				789	39						135.06								
AMMONIA										2.38	0.231	3.150	676.085				0.055		
N_2+Ar										0.05	0.003	0.040	20.921						
H_2										0.04	0.005	0.066					0.041		
H_2S										0.02	0.003	0.051					0.019		
CO_2										0.17	0.006	0.084	5.846				0.198		
HYDROCARBONS										0.84	0.041	0.574	3.736				0.894		
CO										1972.75	561.69	442.87					8.398		
ASH										0.05	0.005	0.004					0.051		
PHENOL											147.25								
NAPHTHA													2.189						
																			265.31
TOTALS				2 694	1 944	14 775	15 485	376	189	1 976.34	844.30	446.92	708.69	11 573	412.72	621.52	9.605	199	265.31
t/d				64 666	46 652	354 602	371 641	9 028	4 546	47 432	20 263	10 726	17 009	277 746	9 905	14 916	231	4 777	6 367
m^3/d																			
SP.GR																			
kg/m^3																			
Sm^3/d																			
TJ/d																			
Note 1					21	27	27	2	2	2	2	2	2	2	2	–	2	2	2

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above
SCF conditions - 60° F, 14.7 PSIA; Sm^3 conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
	82-01-[illegible]	ORIGINAL	[illegible]		

COMPONENT MATERIAL BALANCE

Client EMR
Plant COAL LIQUEFACTION
Project No 1058

DYNAWEST PROJECTS LTD

TABLE 4.1

Sheet 1 of 10

FORM NO. M-01-1081

NUMBER / STREAM / COMPONENT	MOL WT	SG	°API		215 TURBINE FUEL TO HYDRO-PROCESSING	216 GAS OIL TO HYDRO-PROCESSING	217 SOUR WATER FROM HYDROPROCESSING	218 SOUR WATER FROM [illegible]	219 TOTAL SOUR WATER [illegible] SYSTEM	220 VACUUM BOTTOMS TO HYDROGEN PLANT	221 RECYCLE OIL FOR SLURRYING	222 QUENCH OIL FROM FRACTIONATION	223 FUEL GAS FROM AMMONIA HYDRO-PROCESSING	224 AMMONIA PRODUCT TO STORAGE	225 SOUR WATER FROM NAPHTHA HYDRO-PROCESSING	226 SOUR WATER	227 ACID GAS FROM NAPHTHA HYDRO-PROCESSING	228 FUEL GAS FROM [illegible] HYDROPROCESSING	229 FUEL GAS	230 SOUR WATER FROM [illegible] HYDRO-PROCESSING
					m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)	m^3/d (t/h)
TURBINE FUEL					253.88															
GAS OIL						157.05														
AMMONIA							0.063	20.984	22.104			1.383			0.321	22.505				0.679
WATER							93.027	769.112	769.112						3.751	772.845	0.918			3.370
H_2S								5.846	6.096							5.846	0.057			
CO_2								3.736	3.736							3.736	0.278			
PHENOL								2.109	2.109							2.109				
HYDROCARBONS												716.998	0.266					0.378	0.644	
NAPHTHA														265.09						
N_2+Ar																	0.392			
CO																	0.258			
TOTALS					253.88	157.05	93.09	801.787	802.984	597.505	1260.28	718.381	0.266	265.09	4.052	807.038	1.903	0.378	0.644	4.648
t/d					6 093	3 769	2 234	19 243	19 272	14 340	30 247	17 241	6.38	6 361	97.24	19 369	45.68	201.06	207.44	111.66
m^3/d																				
SP GR																				
kg/m^3																				
Sm^3/d																				
TJ/d																				
Note 1					2	2	2			2	2	2	2	2	2		2	2		2

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SM^3 conditions - 60° F, 14.7 PSIA, m^3 conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'KD	APP'D
	82-04-28	ORIGINAL	SHHL		

COMPONENT MATERIAL BALANCE

Client EMR

Plant COAL LIQUEFACTION

Project No 1158

DYNAWEST PROJECTS LTD.

TABLE 4.1

Sheet 2 of 10

FORM NO. H-01-1081

NUMBER	COMPONENT PROPERTIES				231	232	233	234	235	236	237	238	239	240	241	242	243	244	245	246
[illegible]					SOUR WATER	TURBINE FUEL PROD. TO STORAGE	ACID GAS FROM [illegible] HYDRO-PROCESSING	FUEL GAS FROM GAS OIL HYDRO-PROCESSING	FUEL GAS TO UTILITIES	SOUR WATER FROM GAS OIL HYDRO-PROCESSING	SOUR WATER TO SOUR WATER STRIPPING	GAS OIL PROD. TO STORAGE	ACID GAS FROM GAS OIL HYDRO-PROCESSING	ACID GAS	ACID GAS	ACID GAS TO [illegible] SYSTEM	ACID GAS TO [illegible]	CONDENSATE FROM COAL LIQUEFACTION AREA	STEAM FROM COAL LIQUEFACTION AREA	STEAM TO [illegible]
COMPONENT	MOL WT	S.G.	°API		m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]	m³/d [✓ t/h]
AMMONIA					23.184					0.876	24.060									
WATER					776.813					3.005	779.818				2.310	2.810	5.987	308.52	386.21	84.963
H_2S					5.846		0.081				5.846		0.051	0.132	0.642	0.699	1.664			
CO_2					3.736		0.766				3.736		0.797	1.563	244.576	244.854	313.334			
PHENOL					2.109						2.109									
TURBINE FUEL						251.34														
N_2+Ar							0.919						0.956	1.875	1.875	2.267	2.267			
HYDROCARBONS							2.451	10.062	18.696				2.550	5.001	5.129	6.097	8.776			
CO							0.612						0.639	1.251	1.294	1.552	1.552			
GAS OIL												154.07								
H_2															0.022	0.022	0.022			
TOTALS					811.686	251.34	4.829	10.062	18.696	3.881	815.567	154.07	4.993	9.822	255.819	267.762	333.593	308.52	386.21	84.963
t/d					19 480	6 032	115.89	241.49	448.68	93.16	19 574	3 698	119.84	235.73	6 140	6186	8 006	7 405	9 269	2 039
m³/d																				
SP GR																				
kg/m³																				
Sm³/d																				
TJ/d																				
Note 1						2	2	2		2		2	2			2				

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SCF conditions: 60° F, 14.7 PSIA, m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
	82-01-30	ORIGINAL	DWC		

COMPONENT MATERIAL BALANCE

Client: EMR

Plant: COAL LIQUEFACTION

Project No. 1058

DYNAWEST PROJECTS LTD.

TABLE 4.1

Sheet 3 of 10

FORM NO. M-01-1081

NUMBER STREAM COMPONENT	MOL WT	S G	°API	247 STEAM TO HYDROGEN PLANT	248 STEAM FROM COAL [illegible]	301 ACID GAS FROM AMINE PLANT	302 FUEL GAS FROM [illegible] PLANT	303 DRY GAS TO CRYOGENIC PLANT	304 FUEL GAS FROM CRYOGENIC PLANT	305 FUEL GAS TO GAS [illegible]	306 [illegible] PLANT H2 TO H-COAL REACTOR	307 GAS FEED TO DEA AMINE PLANT	309 ACID GAS FROM AMINE PLANT	310 [illegible] GAS TO S.R. PLANT	311 FUEL GAS FROM DEA AMINE PLANT	312 FUEL GAS TO REFORMER	313 FUEL GAS TO [illegible]	314 FUEL GAS TO [illegible]	315 FUEL GAS TO [illegible]
				m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)
H_2O				301.72	429.24	1.636													
H_2S						0.274						4.294	3.677	3.677	0.627	0.559	0.068	0.559	
CO_2						36.194						0.682	0.682	0.956					
H_2												32.396	32.346	68.480					
N_2									4.077	0.816		3.303			3.303	3.478	0.426	7.555	2.581
CO									8.295	1.663		5.269			5.269	4.693	0.576	12.988	6.264
HYDROCARBON									14.118	2.831		6.650			6.650	5.924	0.726	20.891	8.941
									181.438	36.373		145.21	1.033	2.729	144.177	128.436	15.739	309.026	113.892
TOTALS				301.72	429.24	38.104	338.06	338.06	207.928	41.682	88.453	199	37.738	75.842	160.626	143.09	17.535	351.019	131.678
t/d				7 241	10 302	914.49	8 114	8 114	4 990	1 000	2 123	4 777	905.71	1820	3 855	3 434	420.85	8 424	3 160
m³/d																			
SP GR																			
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1						2	2	2	2			2	2	2	2	1	1	1	1

REMARKS

BASIS FOR FLOWS:

Note 1: No. of parallel streams for total flows above

SP conditions: 60°F, 14.7 PSIA; m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	DWN	PREP'D	CH'K'D	APP'D
0	81-01-[illegible]	ORIGINAL				

COMPONENT MATERIAL BALANCE

Client EMR

Plant COAL LIQUEFACTION

Project No. 1058

DYNAWEST PROJECTS LTD.

TABLE 4.1

Sheet 4 of 10

FORM NO. M-01-10 81

NUMBER	COMPONENT PROPERTIES			316	317	318	319	320	321	401	402	403	404	405	406	407	408	409	410
STREAM				FUEL GAS TO COMPRESSOR	FUEL GAS TO [illegible]	NATURAL GAS SUPPLY	FUEL GAS TO [illegible]	FUEL GAS TO COMBUSTOR	PLANT FUEL TO [illegible]	VENT CO₂ GAS TO ATMOS	SOUR WATER FROM GASIFIER	ACID GAS FROM GASIFIER	ASH FROM GASIFIER	HYDROGEN FROM GASIFIER	HYDROGEN TO H-COAL REACTOR	SHIFT GAS TO CO₂ REMOVAL	VENT GAS, CO₂ REMOVAL	HYDROGEN FROM [illegible]	HYDROGEN TO COAL LIQUEFACTION AREA
COMPONENT	MOL WT	S.G.	°API	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h
H_2				0.653	0.653		0.653	1.470	4.221	0.123		0.022		31.980	27.205	64.259	3.357	60.902	85.225
$N_2 + Ar$				0.878	0.878		0.878	2.541	6.856	0.128				16.910	19.340	5.587	0.290	5.297	3.030
CO				1.111	1.111		1.111	3.341	9.990	0.250		0.043		20.095	40.070	5.728	0.194	3.539	2.025
H_2O				0.104	0.104		0.104	0.109	0.964	1.150		2.310	72.397	0.736	0.736				
HYDROCARBONS				24.076	42.771		42.771	79.144	178.897	0.022		0.128		4.055	11.485	13.969	0.700	13.269	7.350
CH_4						38.609	38.609	38.609											
CO_2										1032.0		243.03		6.023	9.442	385.551	380.200	5.261	3.419
NH_3											1.139								
H_2S												0.611							
SOOT													7.681						
ASH													147.262						
TOTALS				26.825	45.518	38.609	84.121	125.809	132.518	1093.751	1.199	246.027	228.330	157.749	208.798	473.093	384.811	88.263	51.049
t/d				643.79	1 092	926.48	2 019	3 019	4 620	26 250	28.783	5 905	5 480	3 786	5 011	11 354	9 236	2 118	1 225
m³/d																			
SP GR																			
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1												9	9	9				1	

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

S.T.P. conditions 60° F, 14.7 PSIA, m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
	82-01-28	ORIGINAL	DWK		

COMPONENT MATERIAL BALANCE

Client EMR

Plant COAL LIQUEFACTION

Project No. 1058

DYNAWEST PROJECTS LTD

TABLE 4.1

Sheet 5 of 19

FORM NO. M-01-1081

COMPONENT	MOL WT	S.G	°API	411	412	413	414	415	416	417	418	419	420	421	422	423	424	425	426
				HYDROGEN TO HYDRO-[illegible]	[illegible]	HYDROGEN TO [illegible]	HYDROGEN TO [illegible] HYDRO-[illegible]	HYDROGEN TO [illegible] HYDRO-[illegible]	STEAM FROM HYDROGEN PLANT	[illegible] WASTE HEAT BOILERS	[illegible] FROM HYDROGEN PLANT	STEAM TO STEAM TURBINE	STEAM TO DESUPER-HEATER	STEAM TO STEAM TURBINE	STEAM FROM HYDROGEN PLANT	STEAM TO [illegible] AREA	STEAM TO [illegible]	[illegible] FROM H_2 PLANT	[illegible] PLANT
				m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)
H_2				25.677	11.001	14.676	10.567	4.109											
N_2 + Ar				2.267	0.356	1.312	0.919	0.392											
CO				1.509	0.639	0.870	0.612	0.258											
CO_2				1.842	0.797	1.049	0.766	0.278											
HYDROCARBONS				5.019	2.550	3.368	2.951	0.918											
H_2O									58.541	437.16	281.31	744.73	236.22	508.51	308.05	259.52	48.751	290.80	599.33
TOTALS				37.214	15.944	21.271	15.315	5.955	58.541	437.16	281.31	744.73	236.22	508.51	308.05	259.52	48.751	290.80	599.33
t/d				893.14	382.65	510.50	367.57	142.98	1 405	10 492	6 751	17 874	5 669	12 204	7 393	6 229	1 170	6 979	14 384
m³/d																			
SP.GR																			
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1					2	2		2											

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SCF conditions: 60° F, 14.7 PSIA; m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
	R2 09-78	ORIGINAL	[illegible]		

COMPONENT MATERIAL BALANCE
Client ENR
Plant COAL LIQUEFACTION
Project No 1058

DYNAWEST PROJECTS LTD.

TABLE 4.1
Sheet 6 of 10

FORM NO. N-01-1001

NUMBER	COMPONENT PROPERTIES			427	428	501	502	503	504	505	601	602	603	701	702	703	704	705	706
STREAM				BLOWDOWN FROM HYDROGEN PLANT	BFW (HP) TO WASTE HEAT BOILER	AIR TO OXYGEN PLANT	NITROGEN DISCHARGE TO ATMOS	OXYGEN TO GASIFIER	CONDENSATE FROM OXYGEN PLANT	STEAM FROM OXYGEN PLANT	SULPHUR TO STORAGE	STEAM FROM [illegible]	BLOWDOWN FROM [illegible]	ACID GAS TO SULPHUR PLANT	AMMONIA TO STORAGE	STRIPPED WATER TO OIL SEPARATION	WATER TO PHENOL RECOVERY	PHENOL TO STORAGE	[illegible]
COMPONENT	MOL WT	S.G	°API	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)
H_2O				54.902	437.16				89.551	463.42		43.333	64.711	34.523		745.295	746.295	0.516	744.78
N_2						1640	1640												
Ar						28.135	16.687	11.452											
O_2						502.37	52.861	449.513											
SULPHUR											6.996								
H_2S														5.846					
NH_3															24.06				
CO_2														3.736					
PHENOL																2.109	2.109	2.102	
TOTALS				54.902	437.16	2 170.57	1709.41	460.965	89.551	463.42	6.996	43.333	64.711	44.105	24.06	747.405	747.405	2.625	744.78
t/d				1 318	10 492	52 089	41 026	10 788	2 149	11 122	167.91	1 040	1 553	1 059	577.44	17 938	17 938	62.996	17 875
m³/d																			
SP GR																			
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1																			

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SCF conditions: 60° F, 14.7 PSIA; m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'KD	APP'D
	82 09 23	ORIGINAL	DWR		

COMPONENT MATERIAL BALANCE

Client: EMR

Plant: COAL LIQUEFACTION

Project No. 1058

DYNAWEST PROJECTS LTD.

TABLE 4.1

Sheet 7 of 10

FORM NO. M-01-1081

STREAM NUMBER					707	708	709	901	902	903	904	905	906	907	908	909	910	911	912	913
STREAM	COMPONENT PROPERTIES				[illegible] cooling water	[illegible] from effluent system	[illegible] to [illegible]	Raw water intake	[illegible] water to [illegible]	Cooling water make-up	[illegible] water to [illegible] system	Cooling water to cooling tower	Process water make-up	BFW make-up from water treatment	Water treatment blow-down	[illegible] water make-up	Sanitary water to [illegible]	[illegible] water softener	[illegible] (clinker)	[illegible]
COMPONENT	MOL WT	S G	TEMP		m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)
H_2O					744.78	361.69	961.02	3735	1577	2 158	12.56	902.33	469.04	930.32	2.577	179.76	256.57	486.38	1 424	223.31
TOTALS					744.78	361.69	361.02	3735	1577	2 158	1 256	902.33	469.04	930.32	2.517	179.76	256.57	436.33	1 424	223.31
				t/d	17 875	8 680	2 495	89 637	37 839	51 798	30 142	21 666	11 137	22 328	60.41	4 314	6 158	10 472	34 171	5 359
				m³/d																
				SP GR																
				kg/m³																
				Sm³/d																
				TJ/d																
				Note 1																

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

SCF conditions 60° F, 14.7 PSIA; m³ conditions 15°C, at 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'KD	APP'D
	82-06-30	ORIGINAL	DWK		

COMPONENT MATERIAL BALANCE

Client EMR

Plant COAL LIQUEFACTION

Project No 1058

DYNAWEST PROJECTS LTD.

TABLE 4.1 Sheet 8 of 10

FORM NO. M-01-1081

STREAM NO.	MOL WT	S.G.	°API	914	915	916	917	918	919	920	921	922	923	924	925	926	927	928	929
STREAM				WATER SOFTENER BLOW DOWN	[illegible] COOLING TOWER MAKE-UP	[illegible] COOLING TOWER LOSS	[illegible] COOLING TOWER BLOWDOWN	STEAM TO OXYGEN PLANT	[illegible] TO [illegible]	[illegible] TO [illegible] PLANT	CONDENSATE TO DEAERATOR	BFW FROM DEAERATOR	BFW TO WASTE HEAT BOILER	BFW TO DESUPERHEATER	STEAM FROM DESUPERHEATER	L.P. BFW FROM [illegible]	STEAM FROM WASTE HEAT BOILER	L.P. STEAM TO STEAM TURBINE	BFW FROM DEAERATOR
COMPONENT				m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)	m³/d (t/h)
H_2O				57.014	1 859	1 778	80.772	83.551	1 468.42	572.75	985.22	414.53	378.61	35.915	272.13	1616.68	183.06	612.30	1423.41
TOTALS				57.014	1 859	1 778	80.772	83.551	1468.42	572.75	985.22	414.55	378.61	35.915	272.13	1616.68	183.06	612.30	1423.41
t/d				1 368	44 605	42 666	1 939	2 005	35 242	13 746	23 645	9 949	9 087	861.97	6 531	38 800	4 393	14 695	34 107
m³/d																			
SP GR																			
kg/m³																			
Sm³/d																			
TJ/d																			
Note 1																			

REMARKS

BASIS FOR FLOWS:-

Note 1: No. of parallel streams for total flows above

S.T. conditions: 60°F, 14.7 PSIA, m³ conditions 15°C, 101 325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
	82 09 27	ORIGINAL	[illegible]		

COMPONENT MATERIAL BALANCE

Client EMR

Plant COAL LIQUEFACTION

Project No 1058

DYNAWEST PROJECTS LTD.

TABLE 4.1

Sheet 9 of 10

FORM NO. M-01-10.81

NUMBER	COMPONENT PROPERTIES				930	931	932	933	934											
STREAM					[illegible]	[illegible] O_2 PLANT (COAL [illegible]	[illegible] OXYGEN PLANT	[illegible] COAL [illegible]	LP [illegible] BOILER											
COMPONENT	MOL WT	S.G.	[illegible]		m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h	m³/d t/h
H_2O					105.53	1328.08	463.47	864.66	183.06											
	TOTALS				105.63	1328.08	463.47	864.66	183.06											
	t/d				2 535	31 874	11 122	20 752	4 393											
	m³/d																			
	SP GR																			
	kg/m³																			
	Sm³/d																			
	TJ/d																			
	Note 1																			

REMARKS

BASIS FOR FLOWS:

Note 1: No. of parallel streams for total flows above

S.T. conditions - 60° F, 14.7 PSIA, m³ conditions 15°C, 101.325 kPa

REV	DATE	DESCRIPTION OF ISSUE	PREP'D	CH'K'D	APP'D
	82 09 22	ORIGINAL	DWL		

COMPONENT MATERIAL BALANCE

Client: EMR

Plant: COAL LIQUEFACTION

Project No: 1058

DYNAWEST PROJECTS LTD.

TABLE 4.1

Sheet 10 of 10

FORM NO. N-01-1081

Table 4.2.1 Major Equipment Schedule for 19 Million t/a Coal Mine, Surface Strip Mining)

Draglines	53 m^3 capacity	4
	44 m^3 capacity	5
Electric Shovels	11.5 m^3 capacity	7
Front End Loaders	9.2 m^3 capacity	3
100 t dump trucks		36

Table 4.6 H-Coal Process Major Utilities Summary

	Power (MW)		Steam (t/sd)		Cooling Water (GJ/h)	
	Produced	Consumed	Produced	Consumed	Produced	Consumed
Coal Preparation	-	47.8	-	-	-	7.5
Coal Liquefaction Fractionation, Hydrotreating	-	155	29 246	7 555	-	115
Light Ends Processing	-	23.3	-	3 905	-	75
Hydrogen Plant Gasification	-	117.3	10 150	6 776	-	980
Hydrogen Plant Reforming	-	125.7	10 644	766	-	365
Oxygen Plant	-	0.8	11 122	2 149	-	1 030
Emission Control Systems	-	2.9	1 040	46	-	-
Effluent Control Systems	-	3.0	-	9 083	-	1 240
Utilities and Offsites	545	67.4	(13 696)	45 620	9 350	4 430
TOTAL	545	545	75 900	75 900	9 350	8 245

Note: Excess capacity is provided in the cooling system.

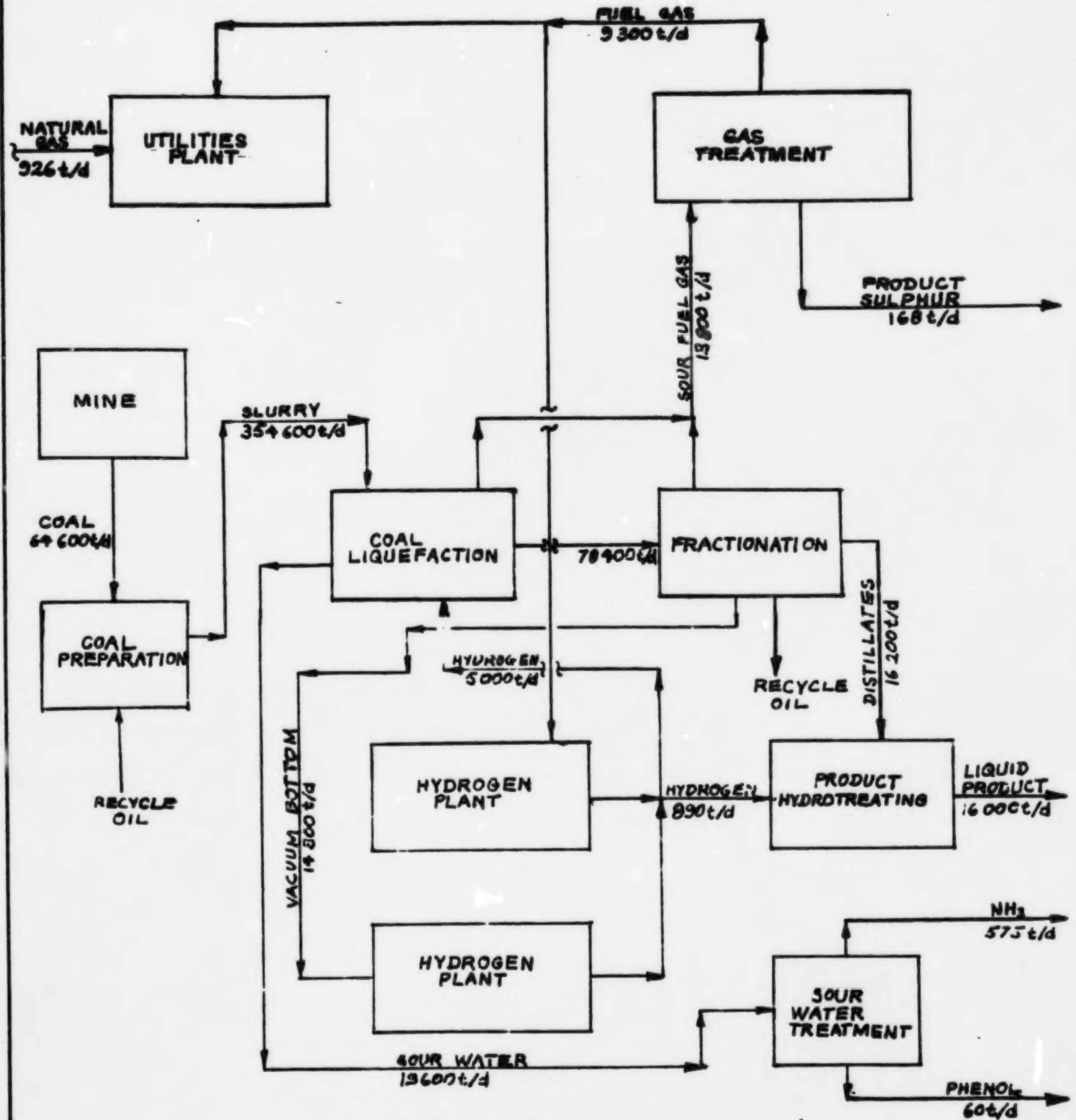
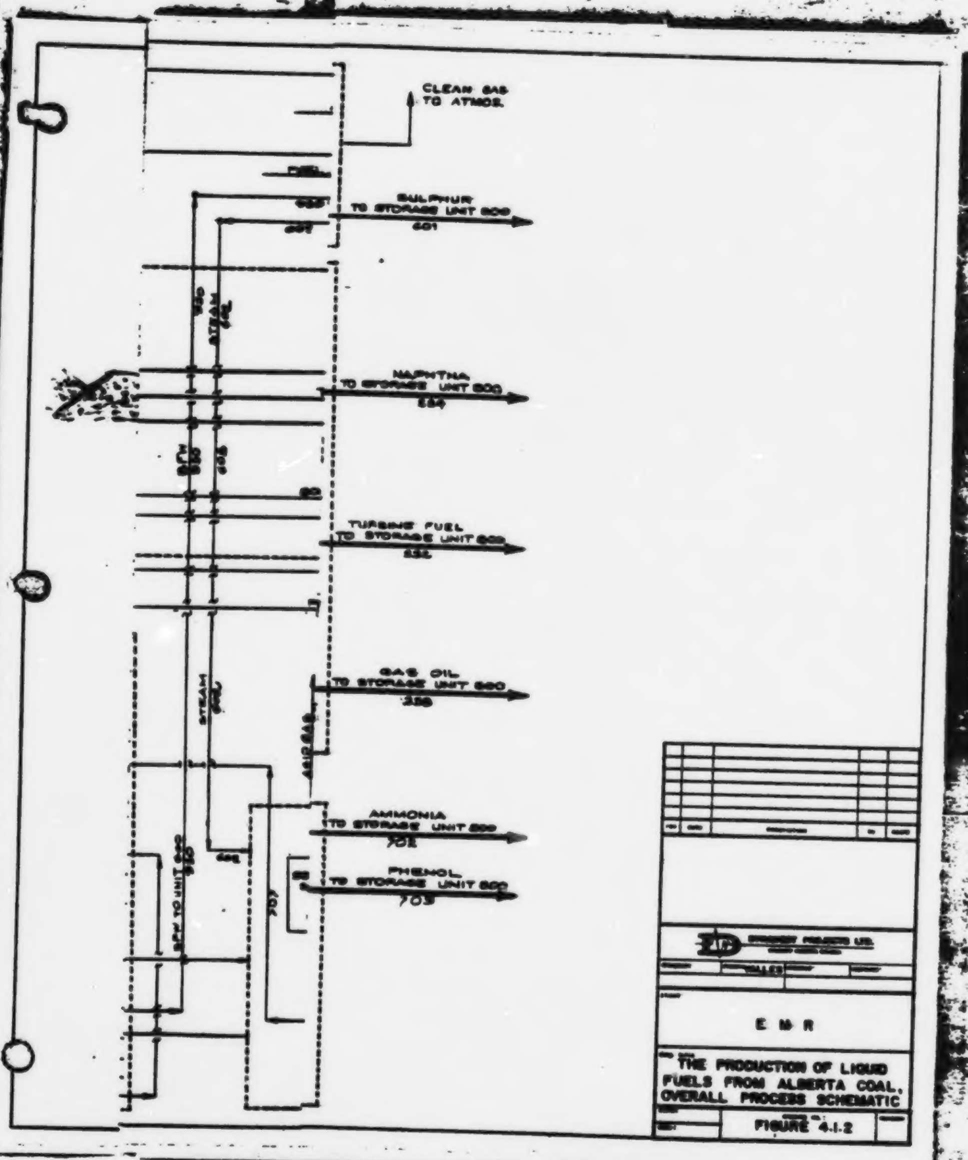

FUEL GAS
9300 t/d
NATURAL GAS
926 t/d
UTILITIES PLANT
GAS TREATMENT
SOUR FUEL GAS
13800 t/d
PRODUCT SULPHUR
168 t/d
MINE
SLURRY
354600 t/d
COAL
64600 t/d
COAL LIQUEFACTION
78400 t/d
FRACTIONATION
COAL PREPARATION
HYDROGEN
5000 t/d
RECYCLE OIL
DISTILLATES
16200 t/d
RECYCLE OIL
HYDROGEN PLANT
HYDROGEN
890 t/d
PRODUCT HYDROTREATING
LIQUID PRODUCT
16000 t/d
VACUUM BOTTOM
14800 t/d
HYDROGEN PLANT
NH_3
575 t/d
SOUR WATER TREATMENT
SOUR WATER
19600 t/d
PHENOL
60 t/d

FIGURE 4.1.1
LIQUID FUELS FROM COAL
OVERALL BLOCK DIAGRAM

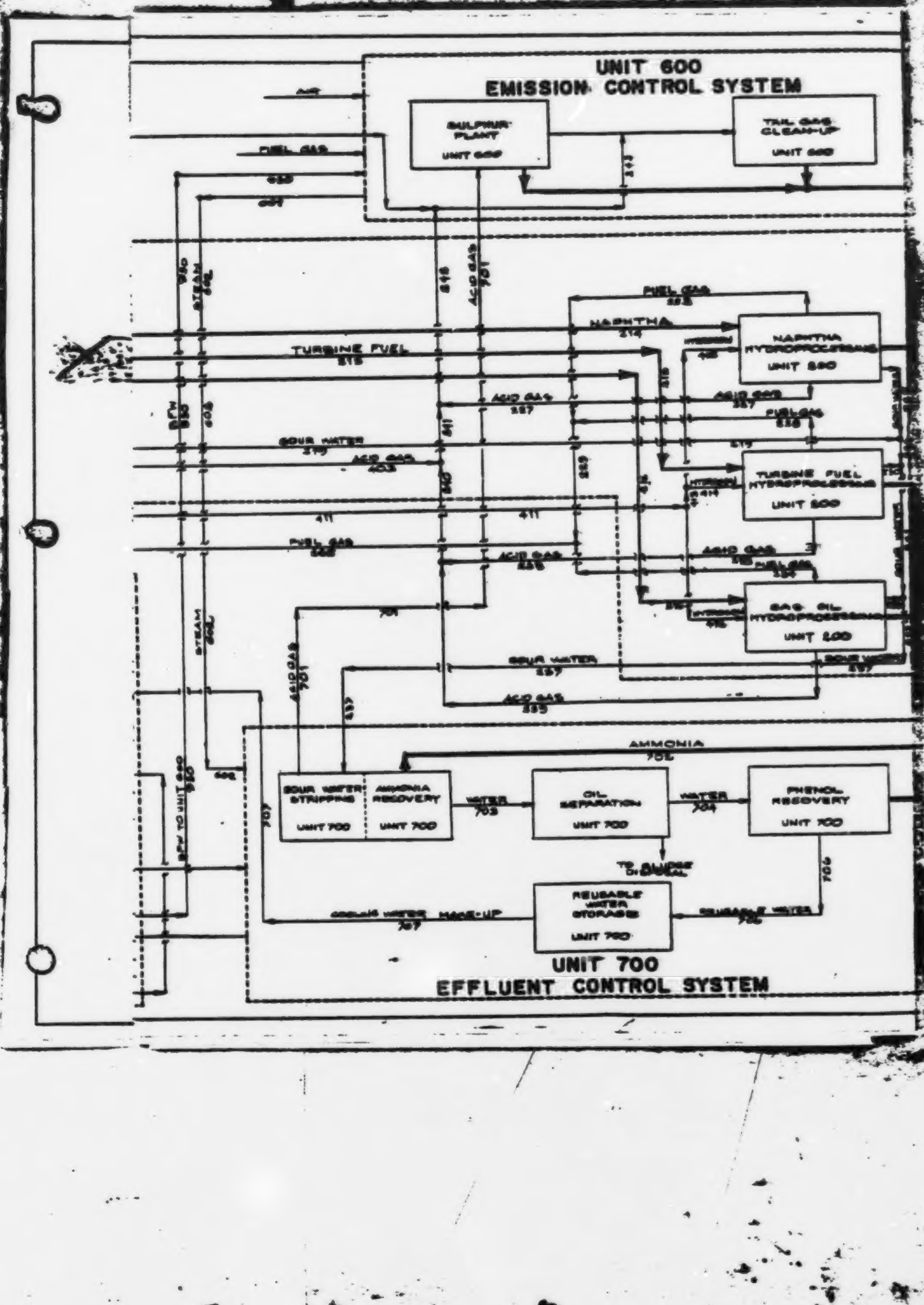
DYNAWEST PROJECTS LTD.

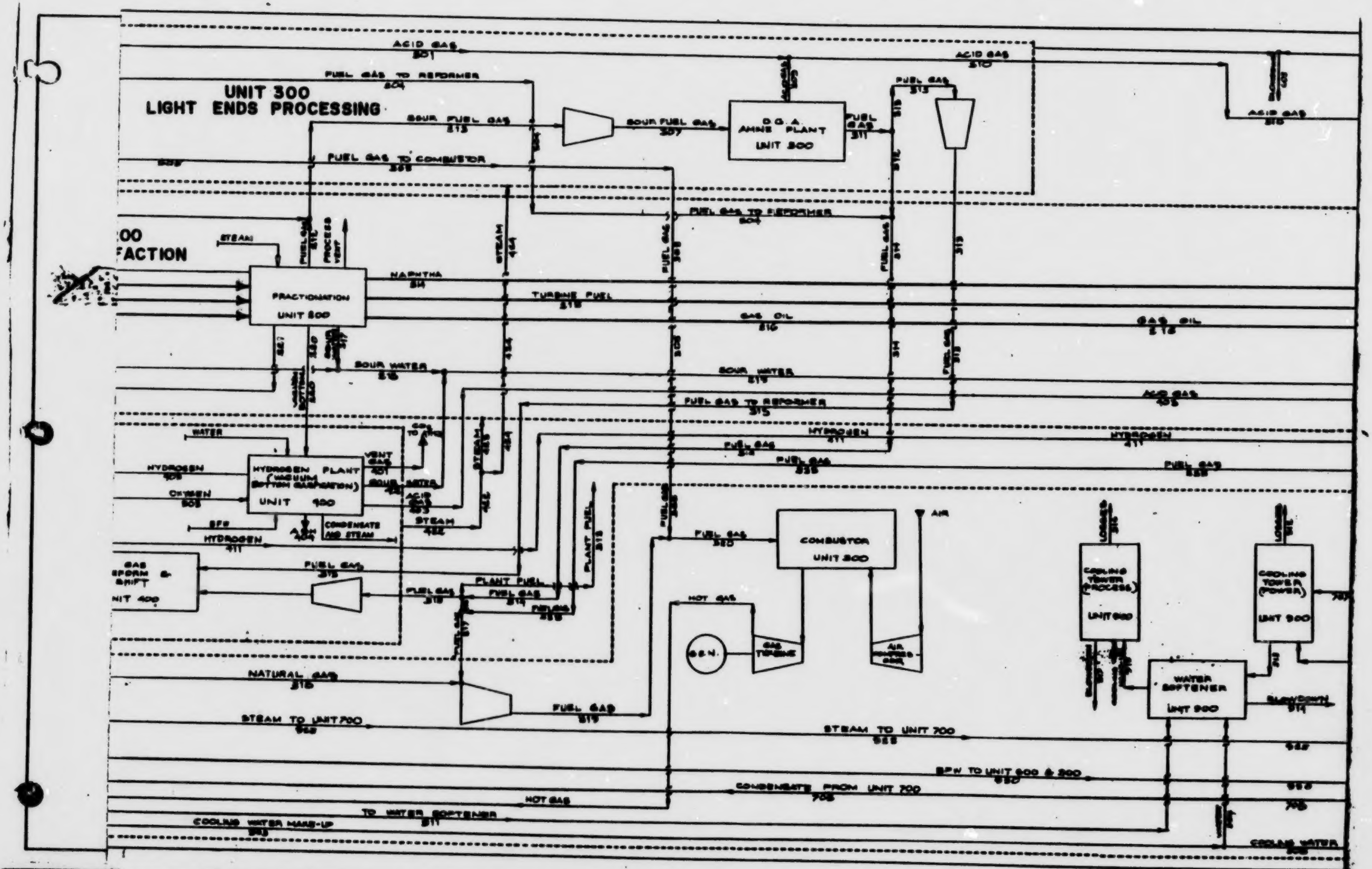
ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET CONT SHT

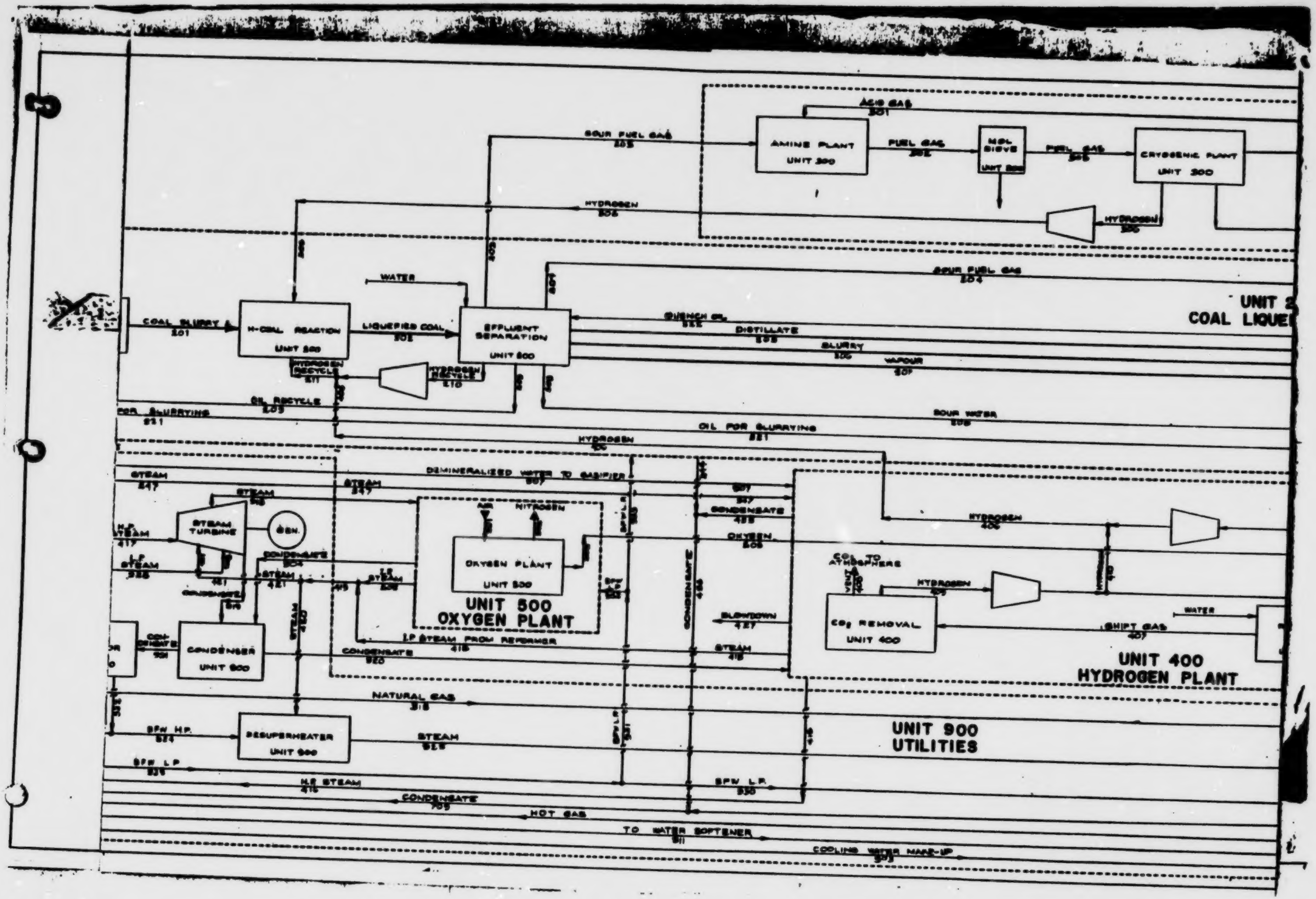

CLEAN GAS TO ATMOS.
SULPHUR TO STORAGE UNIT 800
401
NAPHTHA TO STORAGE UNIT 800
534
TURBINE FUEL TO STORAGE UNIT 800
532
GAS OIL TO STORAGE UNIT 800
ACID GAS
STEAM
AMMONIA TO STORAGE UNIT 800
702
PHENOL TO STORAGE UNIT 800
703
707
BFW TO UNIT 600
E M R

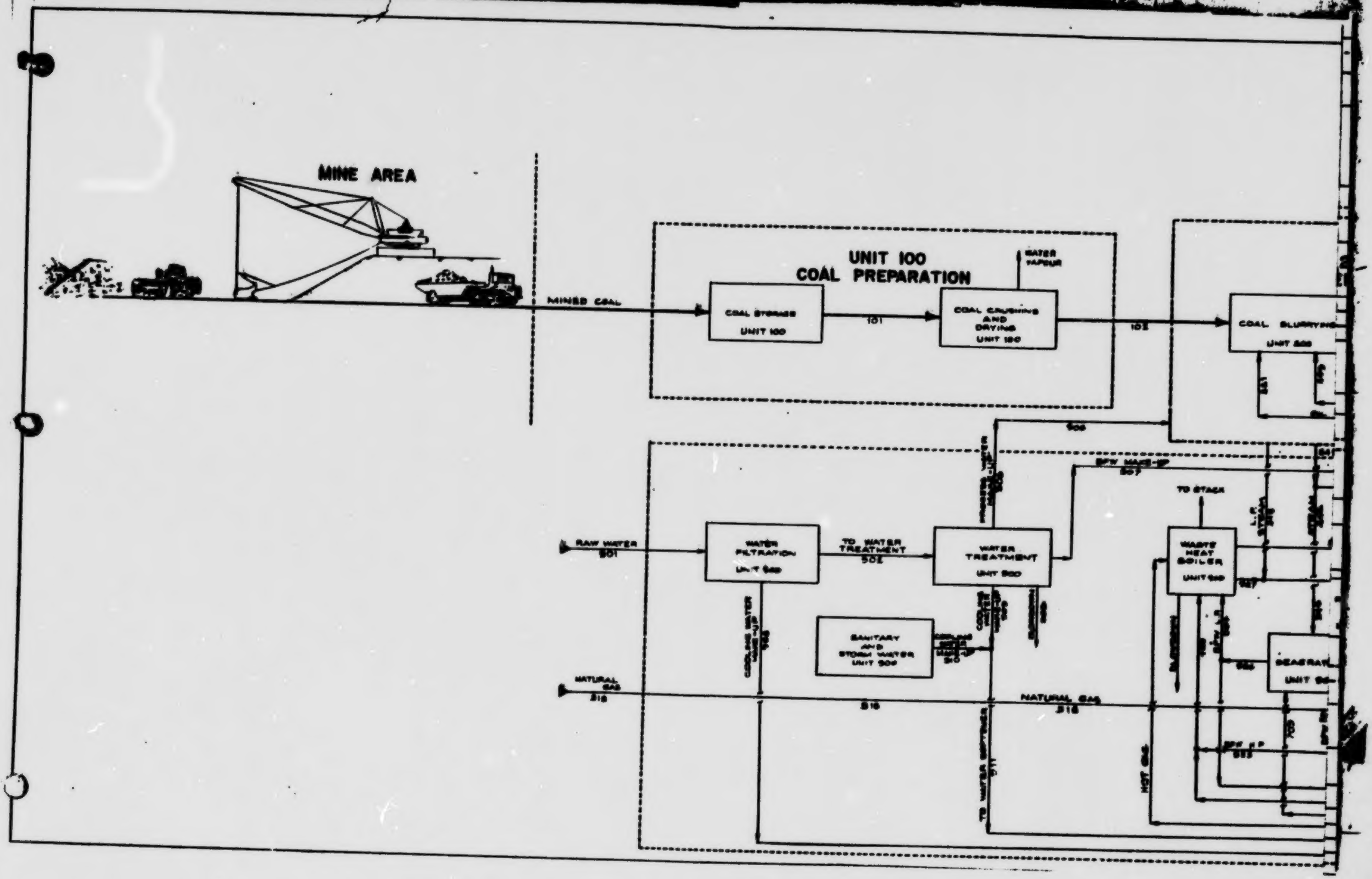
THE PRODUCTION OF LIQUID FUELS FROM ALBERTA COAL, OVERALL PROCESS SCHEMATIC

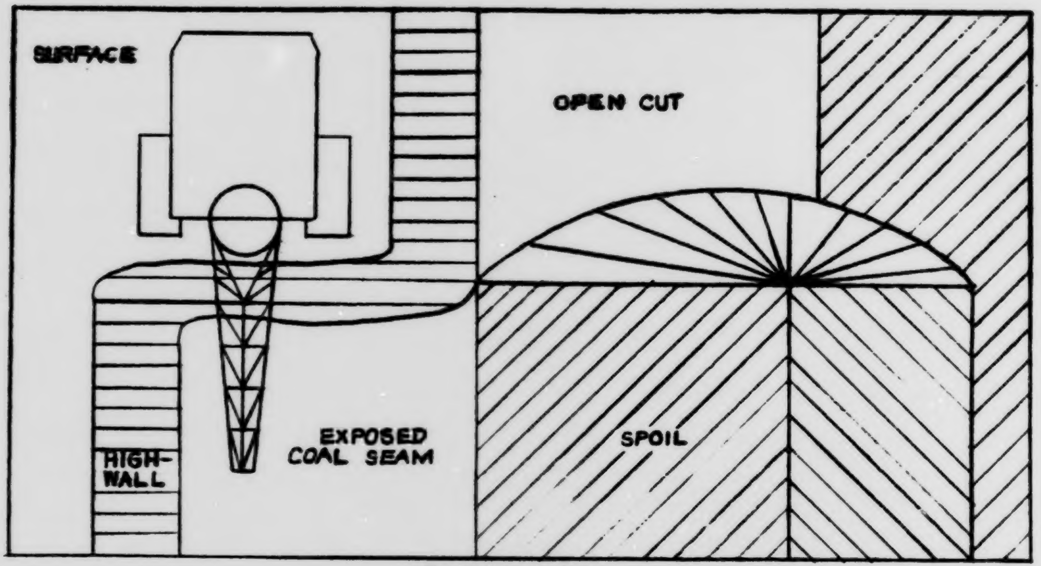
FIGURE 4.1.2


UNIT 600
EMISSION CONTROL SYSTEM
AIR
FUEL GAS
SULPHUR PLANT
UNIT 600
TAIL GAS CLEAN-UP
UNIT 600
STEAM
ACID GAS
701
FUEL GAS
NAPHTHA
214
TURBINE FUEL
215
HYDROGEN
NAPHTHA HYDROPROCESSING
ACID GAS
FUEL GAS
SOUR WATER
ACID GAS
403
411
411
TURBINE FUEL HYDROPROCESSING
UNIT 200
HYDROGEN
FUEL GAS
ACID GAS
ACID GAS
FUEL GAS
GAS OIL HYDROPROCESSING
UNIT 200
HYDROGEN
701
ACID GAS
701
STEAM
SOUR WATER
257
SOUR WATER
257
ACID GAS
255
AMMONIA
702
SOUR WATER STRIPPING
UNIT 700
AMMONIA RECOVERY
UNIT 700
WATER
703
OIL SEPARATION
UNIT 700
WATER
704
PHENOL RECOVERY
UNIT 700
TO SLUDGE DISPOSAL
706
REUSABLE WATER STORAGE
UNIT 700
REUSABLE WATER
COOLING WATER MAKE-UP
707
707
BFW TO UNIT 600
UNIT 700
EFFLUENT CONTROL SYSTEM

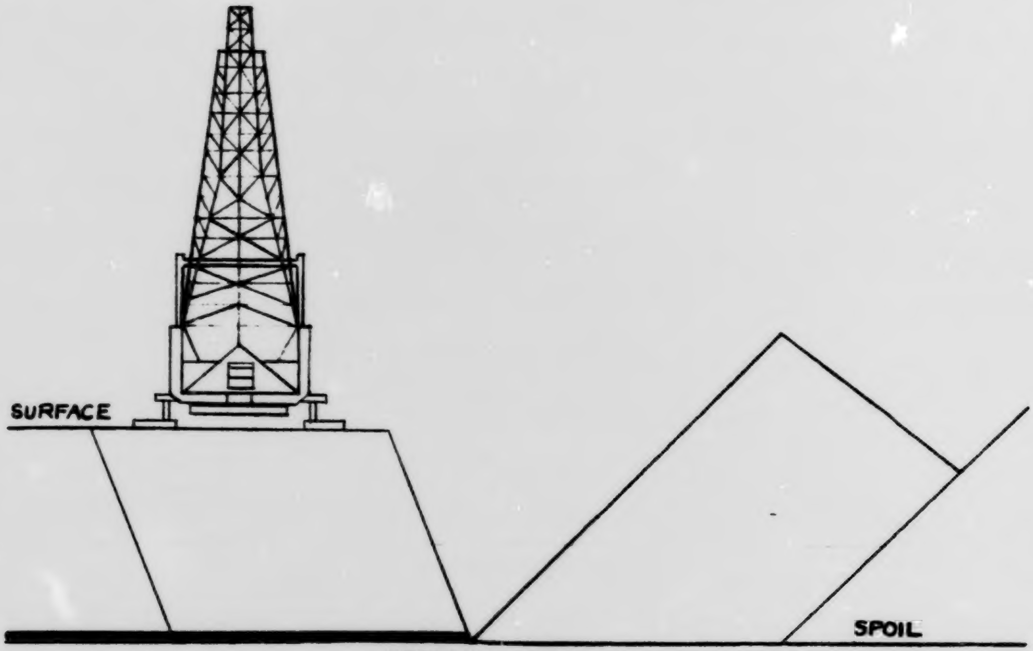

UNIT 300
LIGHT ENDS PROCESSING
00
FACTION
ACID GAS 301
FUEL GAS TO REFORMER 304
SOUR FUEL GAS 313
SOUR FUEL GAS 307
D.G.A. AMINE PLANT UNIT 300
FUEL GAS 311
ACID GAS 305
FUEL GAS 313
ACID GAS 310
BLOWDOWN
ACID GAS 310
FUEL GAS TO COMBUSTOR 308
STEAM
FRACTIONATION UNIT 200
FUEL GAS 312
PROCESS VENT
STEAM 464
FUEL GAS 308
FUEL GAS TO REFORMER 304
FUEL GAS 314
313
NAPHTHA 314
TURBINE FUEL 315
GAS OIL 316
GAS OIL 316
SOUR WATER 316
SOUR WATER 319
VACUUM BOTTOMS 240
SOUR WATER 317
ACID GAS 403
FUEL GAS TO REFORMER 315
WATER
CO2 TO ATM
VENT GAS 401
HYDROGEN 411
HYDROGEN 411
HYDROGEN 405
HYDROGEN PLANT (VACUUM BOTTOM GASIFICATION) UNIT 400
SOUR WATER 402
OXYGEN 503
ACID GAS 403
STEAM 465
STEAM 462
BFW
ASH 404
CONDENSATE AND STEAM
HYDROGEN 411
FUEL GAS 314
FUEL GAS 335
FUEL GAS 335
GAS REFORM & SHIFT UNIT 400
FUEL GAS 318
FUEL GAS 318
PLANT FUEL
FUEL GAS 314
FUEL GAS 335
PLANT FUEL 318
FUEL GAS 317
FUEL GAS 320
COMBUSTOR UNIT 200
AIR
HOT GAS
GEN.
GAS TURBINE
AIR COMPRESSOR
COOLING TOWER (PROCESS) UNIT 900
LOSSES 315
COOLING TOWER (POWER) UNIT 900
LOSSES 912
WATER SOFTENER UNIT 900
BLOWDOWN 914
NATURAL GAS 315
FUEL GAS 319
STEAM TO UNIT 700 962
STEAM TO UNIT 700 965
BFW TO UNIT 900 & 200 950
CONDENSATE FROM UNIT 700 705
HOT GAS
TO WATER SOFTENER 911
COOLING WATER MAKE-UP 913
COOLING WATER 906
962
955
705
707


UNIT 2
COAL LIQUE
UNIT 500
OXYGEN PLANT
UNIT 400
HYDROGEN PLANT
UNIT 900
UTILITIES
COAL SLURRY
201
H-COAL REACTION
UNIT 200
LIQUEFIED COAL
202
EFFLUENT SEPARATION
UNIT 200
WATER
HYDROGEN RECYCLE
211
HYDROGEN RECYCLE
210
OIL RECYCLE
203
OIL FOR SLURRYING
221
SOUR FUEL GAS
203
AMINE PLANT
UNIT 300
ACID GAS
301
FUEL GAS
302
MOL SIEVE
UNIT 300
FUEL GAS
303
CRYOGENIC PLANT
UNIT 300
HYDROGEN
305
HYDROGEN
206
SOUR FUEL GAS
204
QUENCH OIL
222
DISTILLATE
205
SLURRY
206
VAPOUR
207
SOUR WATER
208
OIL FOR SLURRYING
221
HYDROGEN
406
STEAM
347
DEMINERALIZED WATER TO GASIFIER
907
STEAM
347
STEAM
H.P. STEAM
417
L.P. STEAM
328
STEAM TURBINE
GEN.
CONDENSATE
CONDENSER
UNIT 900
STEAM
421
L.P. STEAM
AIR
NITROGEN
OXYGEN PLANT
UNIT 500
L.P. STEAM FROM REFORMER
415
CONDENSATE
920
CONDENSATE
450
CONDENSATE
435
OXYGEN
305
HYDROGEN
406
CO2 TO ATMOSPHERE
HYDROGEN
405
CO2 REMOVAL
UNIT 400
WATER
SHIFT GAS
407
BLOWDOWN
427
STEAM
418
NATURAL GAS
318
DESUPERHEATER
UNIT 900
BFW H.P.
324
STEAM
325
BFW L.P.
325
H.P. STEAM
416
CONDENSATE
705
HOT GAS
BFW L.P.
330
TO WATER SOFTENER
911
COOLING WATER MAKE-UP
903


MINE AREA
MINED COAL
UNIT 100
COAL PREPARATION
COAL STORAGE
UNIT 100
101
COAL CRUSHING
AND
DRYING
UNIT 120
WATER VAPOUR
102
COAL SLURRYING
UNIT 200
RAW WATER
501
WATER FILTRATION
UNIT 540
TO WATER TREATMENT
502
WATER TREATMENT
UNIT 500
COOLING WATER MAKE-UP
PROCESS WATER MAKE-UP
506
BFW MAKE-UP
507
BLOWDOWN
SANITARY
AND
STORM WATER
UNIT 509
COOLING WATER MAKE-UP
510
TO WATER SOFTENER
511
NATURAL GAS
515
516
NATURAL GAS
516
TO STACK
WASTE HEAT BOILER
UNIT 540
BLOWDOWN
BFW L.P.
L.P. STEAM
STEAM
BFW H.P.
DEAERATOR
UNIT 54
HOT GAS


SURFACE
OPEN CUT
HIGH-WALL
EXPOSED COAL SEAM
SPOIL

PLAN VIEW


SURFACE
SPOIL

SECTION VIEW

FIGURE 4.2.1
SINGLE DRAGLINE OPERATION

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

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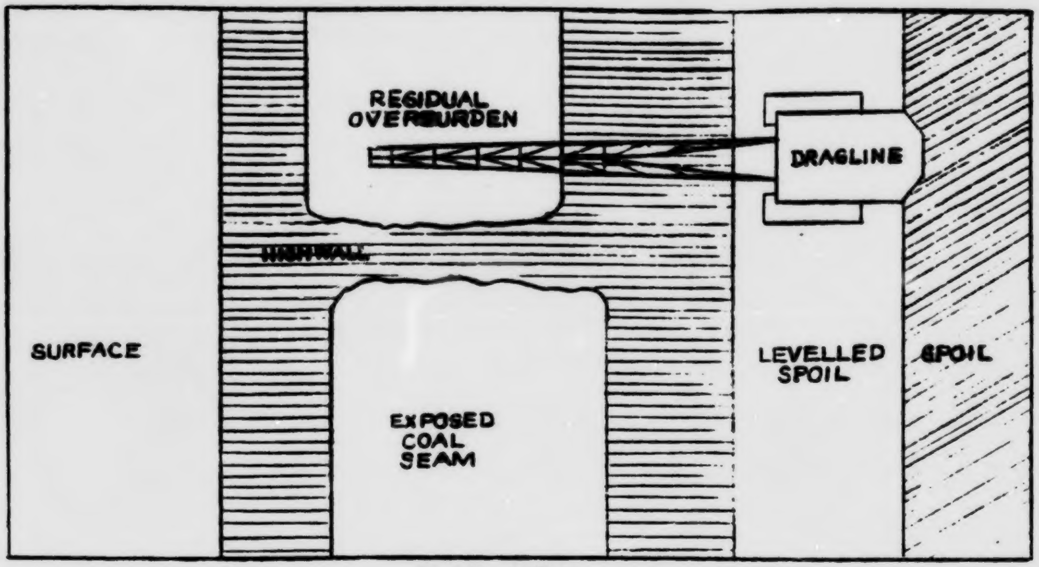
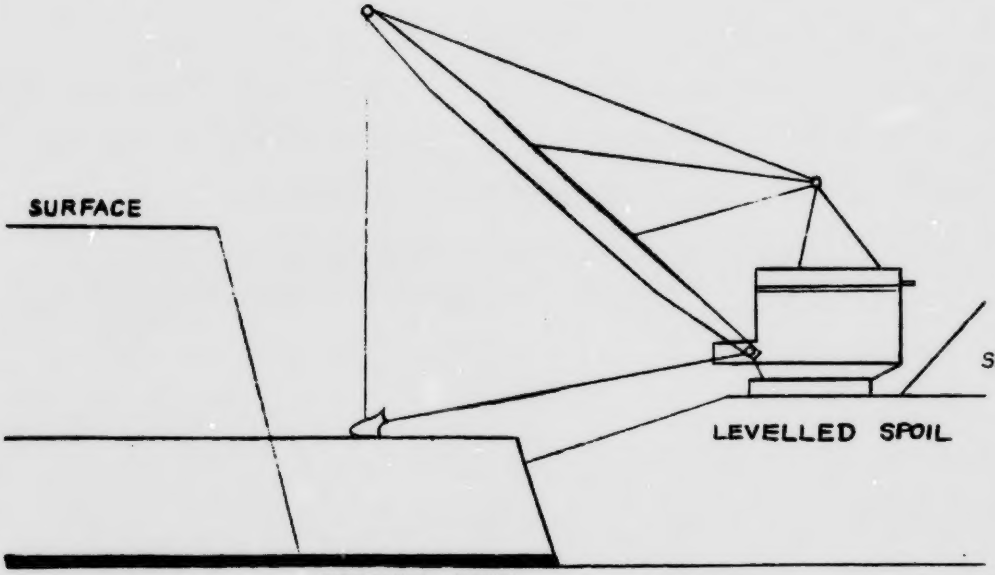
SHEET

CONT SHT


RESIDUAL
OVERBURDEN
DRAGLINE
HIGHWALL
SURFACE
LEVELLED
SPOIL
SPOIL
EXPOSED
COAL
SEAM

PLAN VIEW


SURFACE
SPOIL
LEVELLED SPOIL

SECTION VIEW

FIGURE 4.2.2
TANDEM OPERATION.
POSITION OF SECOND MACHINE

DYNAWEST PROJECTS LTD.

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET
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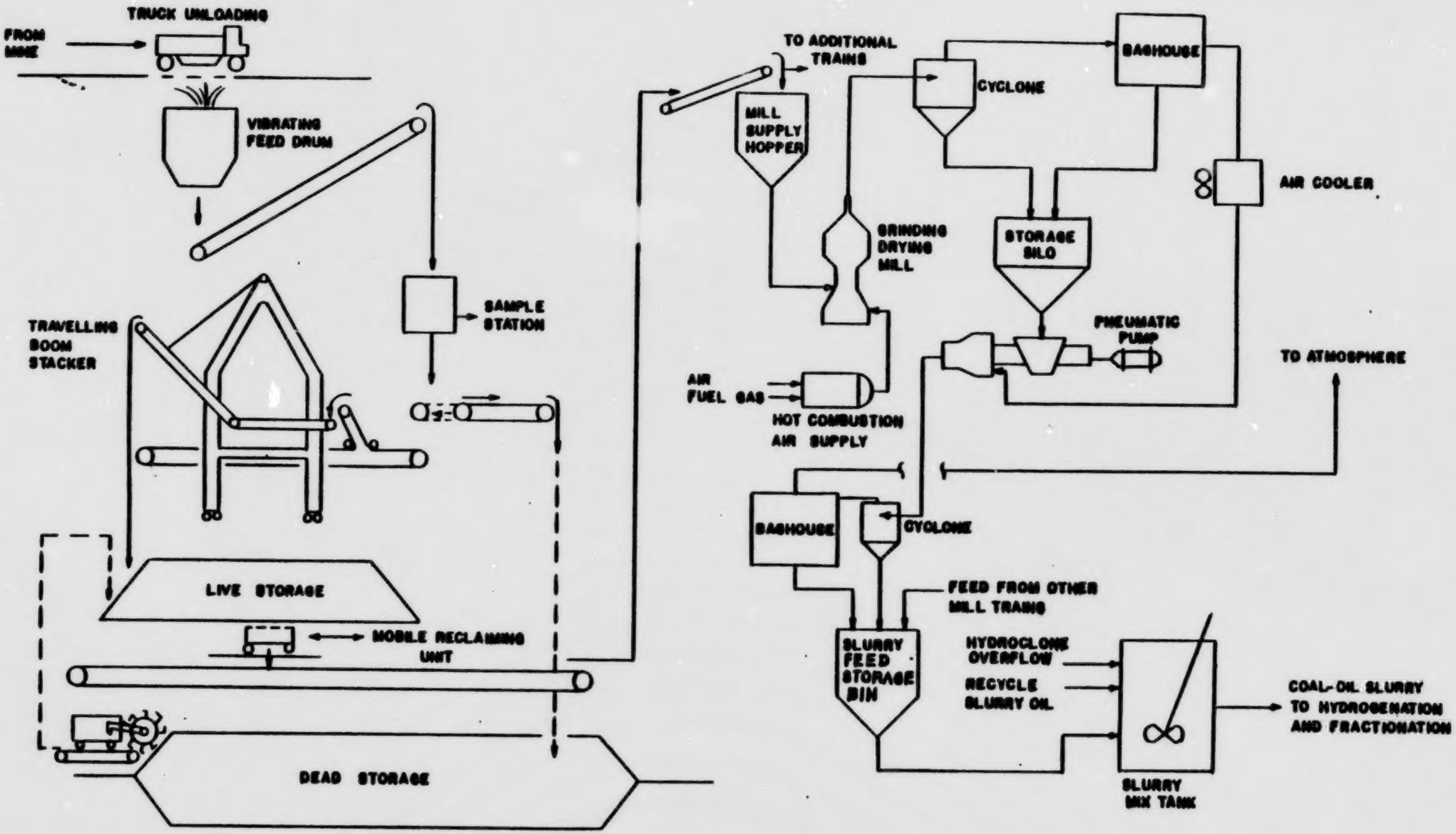
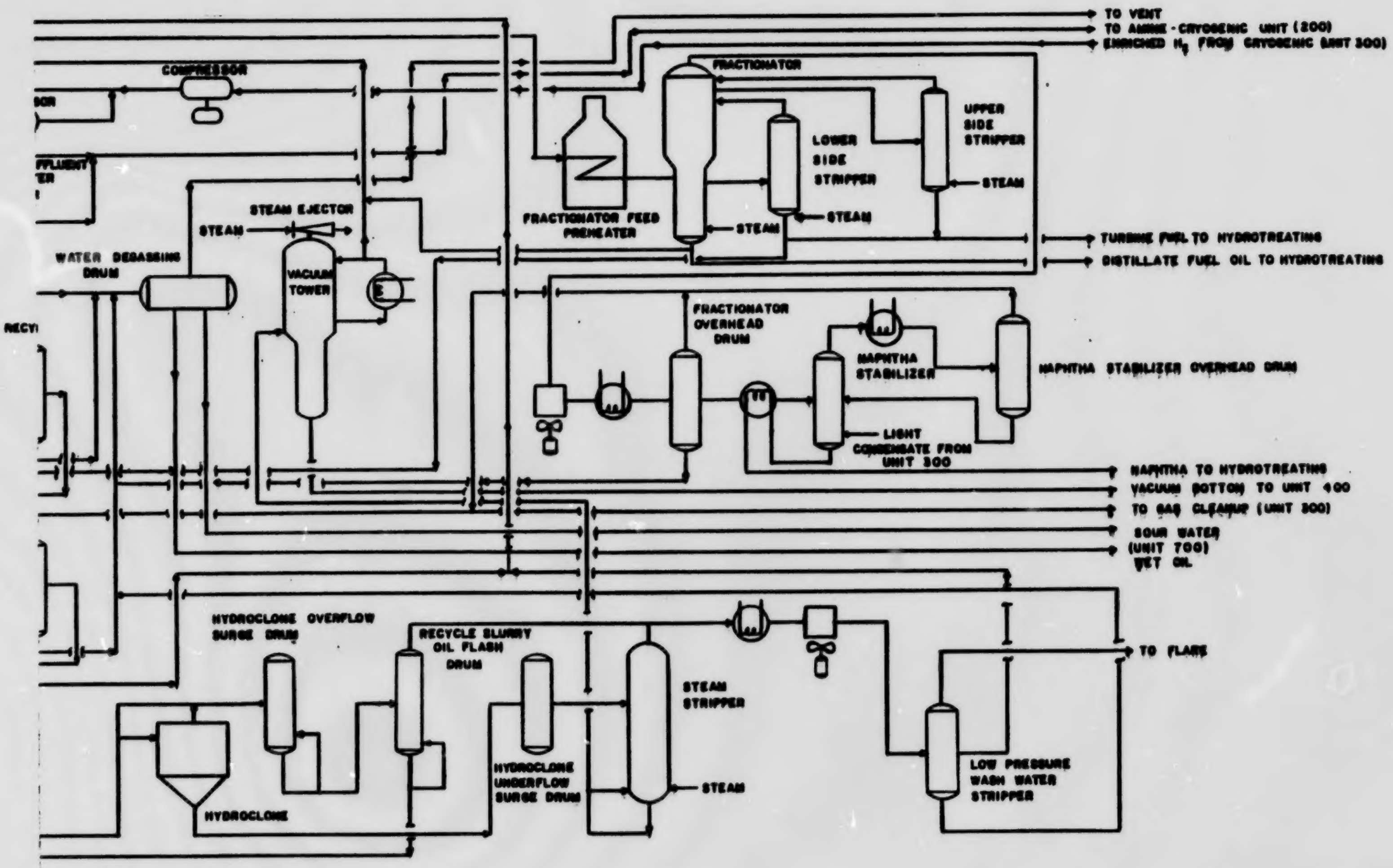

FROM MINE
TRUCK UNLOADING
VIBRATING FEED DRUM
SAMPLE STATION
TRAVELLING BOOM STACKER
LIVE STORAGE
MOBILE RECLAIMING UNIT
DEAD STORAGE
TO ADDITIONAL TRAINS
MILL SUPPLY HOPPER
GRINDING DRYING MILL
AIR
FUEL GAS
HOT COMBUSTION AIR SUPPLY
CYCLONE
BAGHOUSE
STORAGE SILO
AIR COOLER
PNEUMATIC PUMP
TO ATMOSPHERE
BAGHOUSE
CYCLONE
FEED FROM OTHER MILL TRAINS
SLURRY FEED STORAGE BIN
HYDROCLONE OVERFLOW
RECYCLE SLURRY OIL
SLURRY MIX TANK
COAL-OIL SLURRY TO HYDROGENATION AND FRACTIONATION

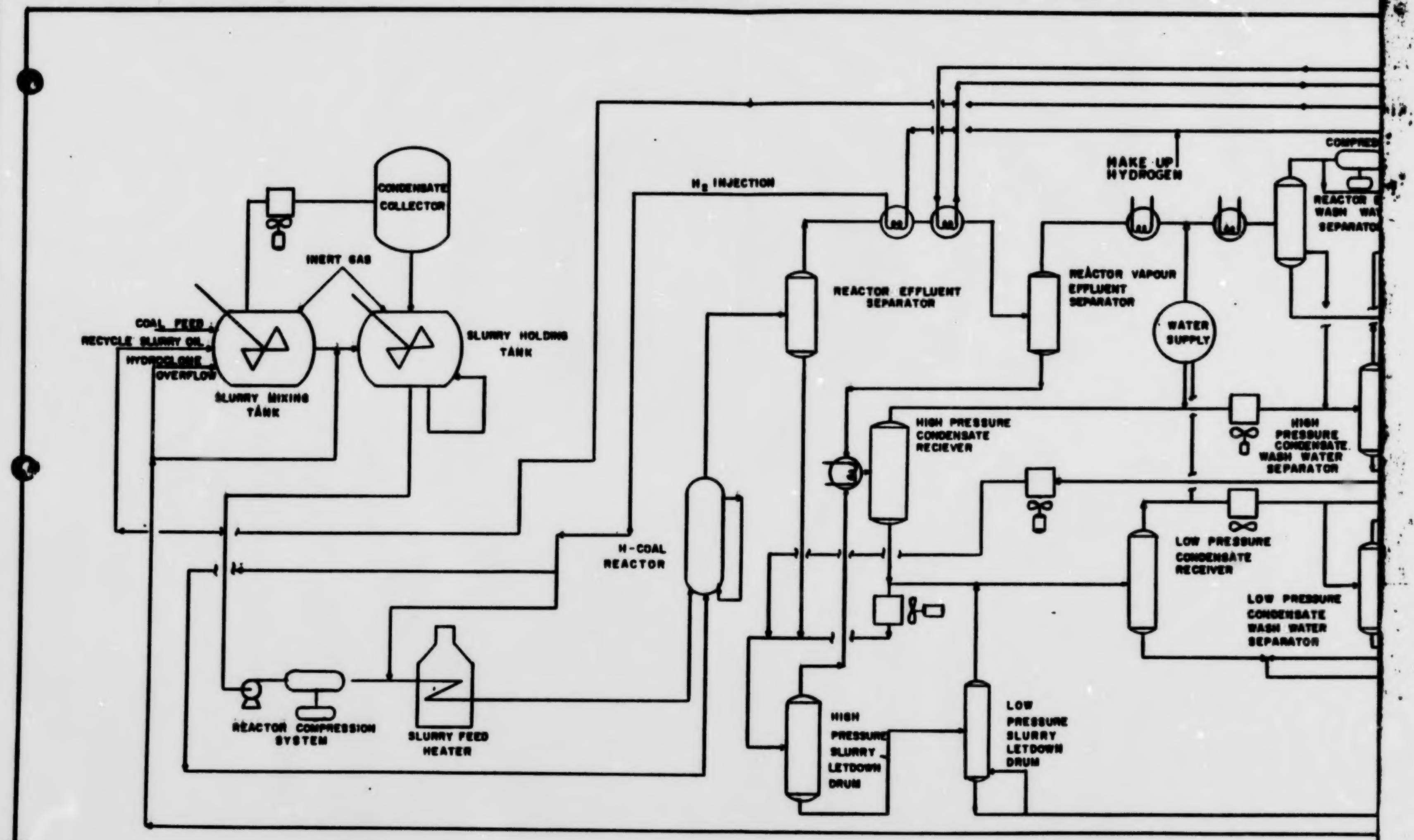
FIGURE 4·3·1
UNIT 100 - COAL STORAGE AND PREPARATION

NOTE: ONLY 1 TRAIN SHOWN

DYNAWEST PROJECTS LTD.


TO VENT
TO AMINE - CRYOGENIC UNIT (200)
ENRICHED H₂ FROM CRYOGENIC (UNIT 300)
COMPRESSOR
SOR
FFLUENT
TER
STEAM EJECTOR
STEAM
WATER DEGASSING DRUM
VACUUM TOWER
RECY
FRACTIONATOR
FRACTIONATOR FEED PREHEATER
LOWER SIDE STRIPPER
UPPER SIDE STRIPPER
STEAM
STEAM
STEAM
TURBINE FUEL TO HYDROTREATING
DISTILLATE FUEL OIL TO HYDROTREATING
FRACTIONATOR OVERHEAD DRUM
NAPHTHA STABILIZER
NAPHTHA STABILIZER OVERHEAD DRUM
LIGHT CONDENSATE FROM UNIT 300
NAPHTHA TO HYDROTREATING
VACUUM BOTTOM TO UNIT 400
TO GAS CLEANUP (UNIT 300)
SOUR WATER
(UNIT 700)
WET OIL
HYDROCLONE OVERFLOW SURGE DRUM
RECYCLE SLURRY OIL FLASH DRUM
HYDROCLONE
HYDROCLONE UNDERFLOW SURGE DRUM
STEAM STRIPPER
STEAM
TO FLARE
LOW PRESSURE WASH WATER STRIPPER

URE 4.4.1
HYDROGENATION AND FRACTIONATION


CONDENSATE COLLECTOR
INERT GAS
COAL FEED
RECYCLE SLURRY OIL
HYDROCLONE OVERFLOW
SLURRY MIXING TANK
SLURRY HOLDING TANK
H_2 INJECTION
H-COAL REACTOR
REACTOR COMPRESSION SYSTEM
SLURRY FEED HEATER
REACTOR EFFLUENT SEPARATOR
REACTOR VAPOUR EFFLUENT SEPARATOR
MAKE UP HYDROGEN
COMPRES
REACTOR
WASH WA
SEPARATO
WATER SUPPLY
HIGH PRESSURE CONDENSATE RECIEVER
HIGH PRESSURE CONDENSATE WASH WATER SEPARATOR
LOW PRESSURE CONDENSATE RECEIVER
LOW PRESSURE CONDENSATE WASH WATER SEPARATOR
HIGH PRESSURE SLURRY LETDOWN DRUM
LOW PRESSURE SLURRY LETDOWN DRUM

FI

COAL LIQUEFACTION UNIT 200-

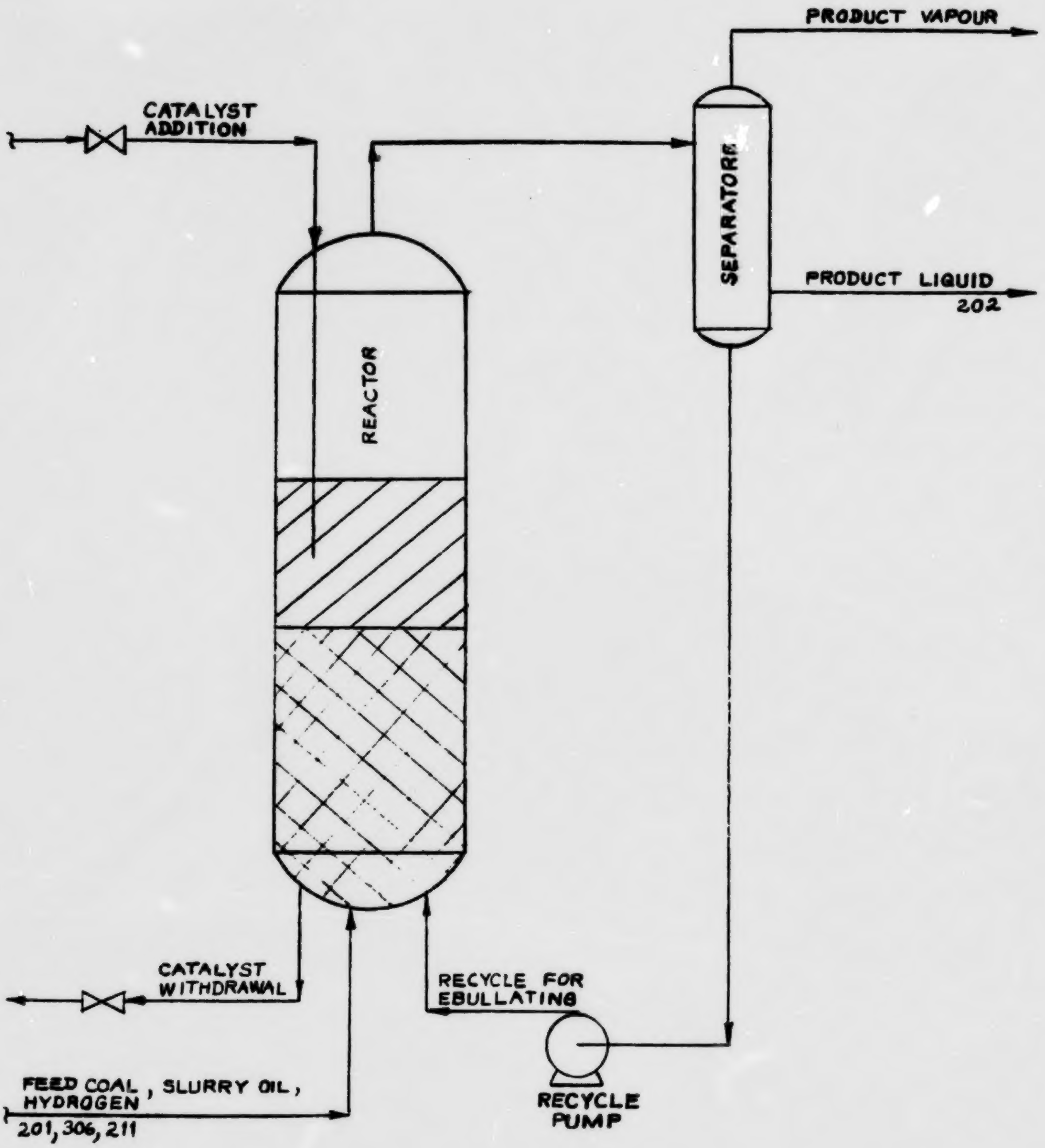

PRODUCT VAPOUR
CATALYST ADDITION
SEPARATORS
PRODUCT LIQUID
202
REACTOR
CATALYST WITHDRAWAL
RECYCLE FOR EBULLATING
FEED COAL, SLURRY OIL, HYDROGEN
201, 306, 211
RECYCLE PUMP

FIGURE 4.4.2
H-COAL REACTOR — UNIT 200

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

DYNAWEST PROJECTS LTD.

SHEET

CONT SHT

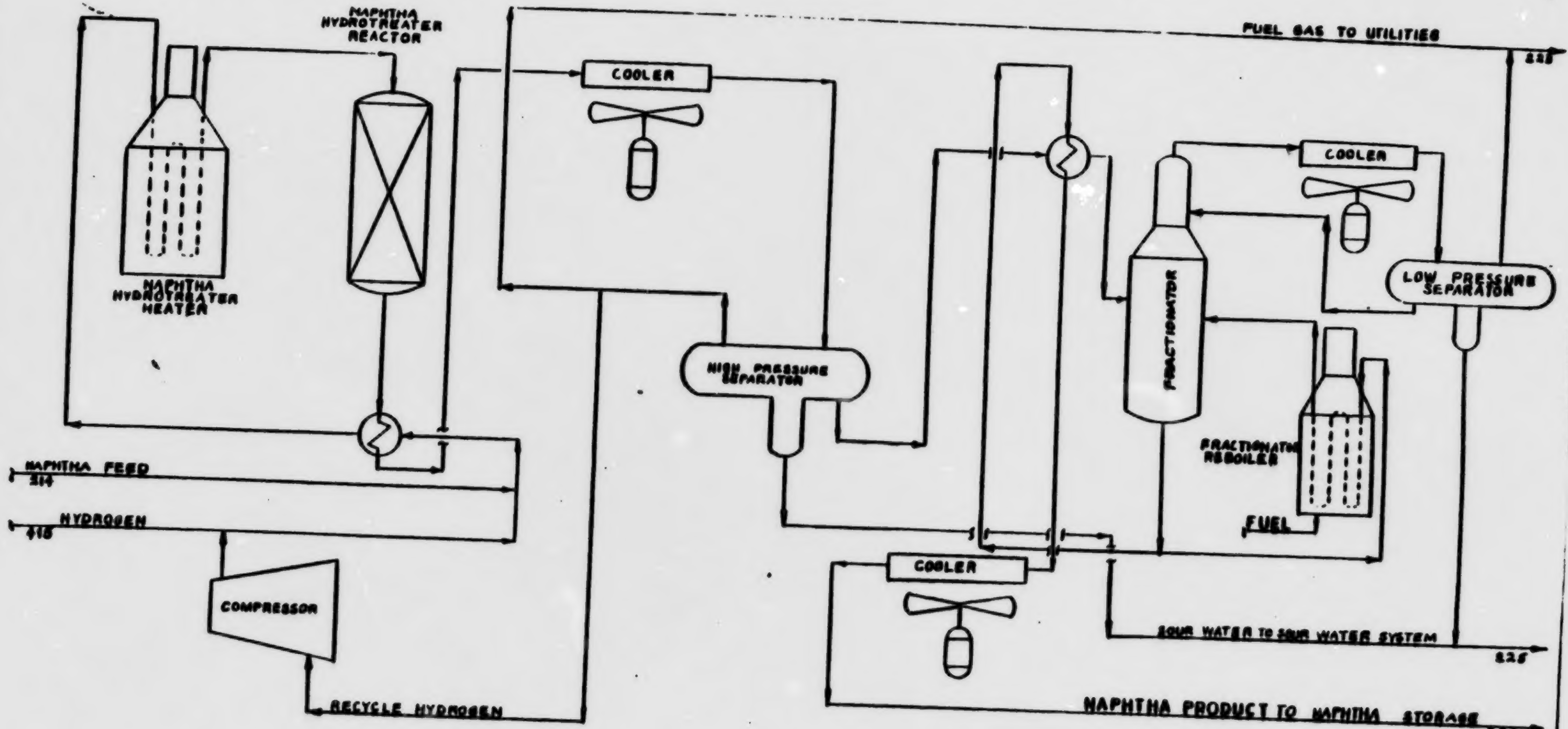
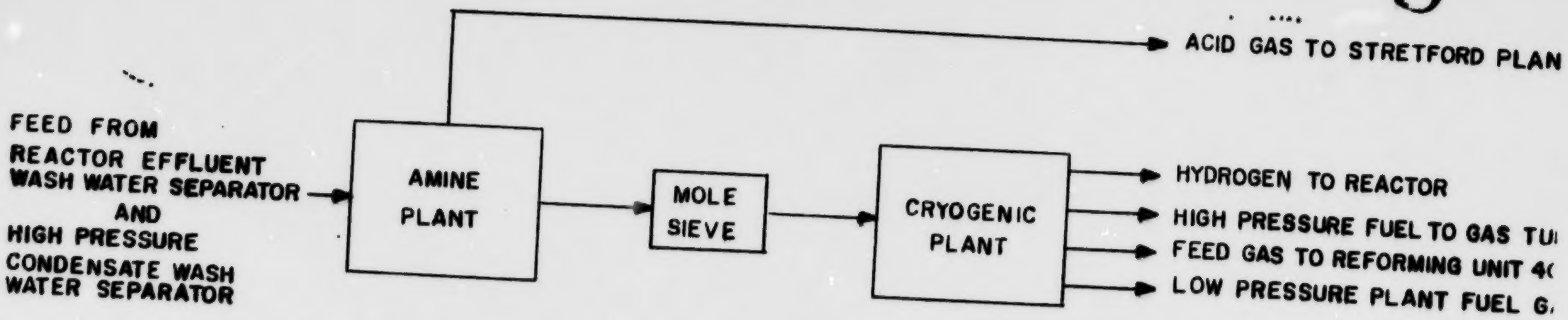

NAPHTHA HYDROTREATER REACTOR
NAPHTHA HYDROTREATER HEATER
NAPHTHA FEED
214
HYDROGEN
418
COMPRESSOR
RECYCLE HYDROGEN
COOLER
HIGH PRESSURE SEPARATOR
COOLER
FRACTIONATOR
COOLER
LOW PRESSURE SEPARATOR
FRACTIONATOR REBOILER
FUEL
FUEL GAS TO UTILITIES
225
SOUR WATER TO SOUR WATER SYSTEM
226
NAPHTHA PRODUCT TO NAPHTHA STORAGE
224

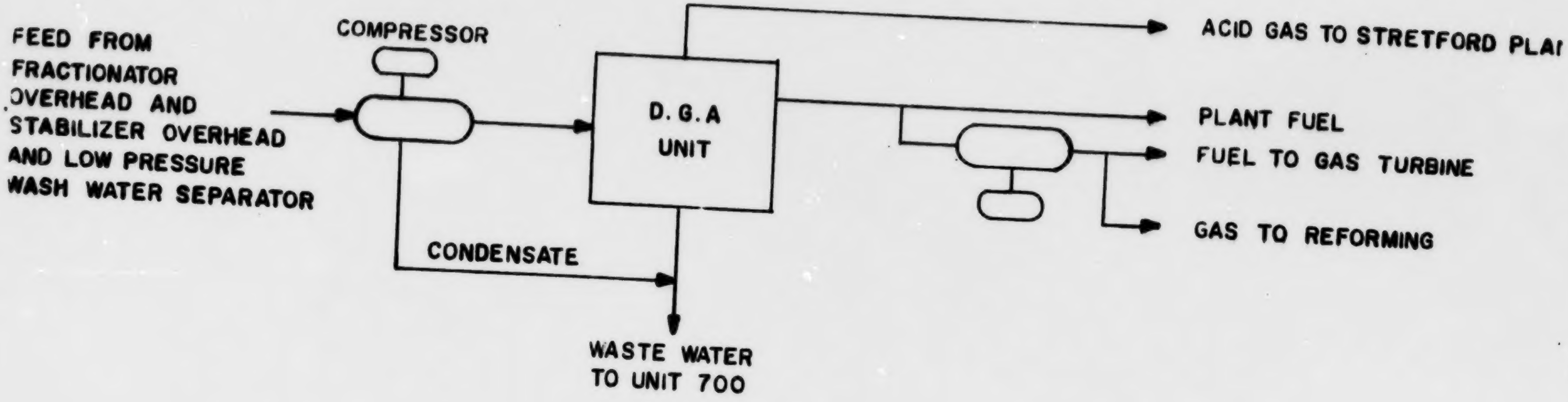
FIGURE 4.4.2.1
NAPHTHA HYDROTREATER

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

DYNAWEST PROJECTS LTD.
SHEET
CONT. SHEET


FEED FROM
REACTOR EFFLUENT
WASH WATER SEPARATOR
AND
HIGH PRESSURE
CONDENSATE WASH
WATER SEPARATOR
AMINE PLANT
MOLE SIEVE
CRYOGENIC PLANT
ACID GAS TO STRETFORD PLAN
HYDROGEN TO REACTOR
HIGH PRESSURE FUEL TO GAS TU
FEED GAS TO REFORMING UNIT 4
LOW PRESSURE PLANT FUEL G

FIRST GAS TREATMENT TRAIN


FEED FROM
FRACTIONATOR
OVERHEAD AND
STABILIZER OVERHEAD
AND LOW PRESSURE
WASH WATER SEPARATOR
COMPRESSOR
CONDENSATE
D. G. A
UNIT
WASTE WATER
TO UNIT 700
ACID GAS TO STRETFORD PLA
PLANT FUEL
FUEL TO GAS TURBINE
GAS TO REFORMING

SECOND GAS TREATMENT TRAIN

FIGURE 4·5·1·1

UNIT 300- LIGHT ENDS PROCESSING

DYNAWEST PROJECTS LTD.

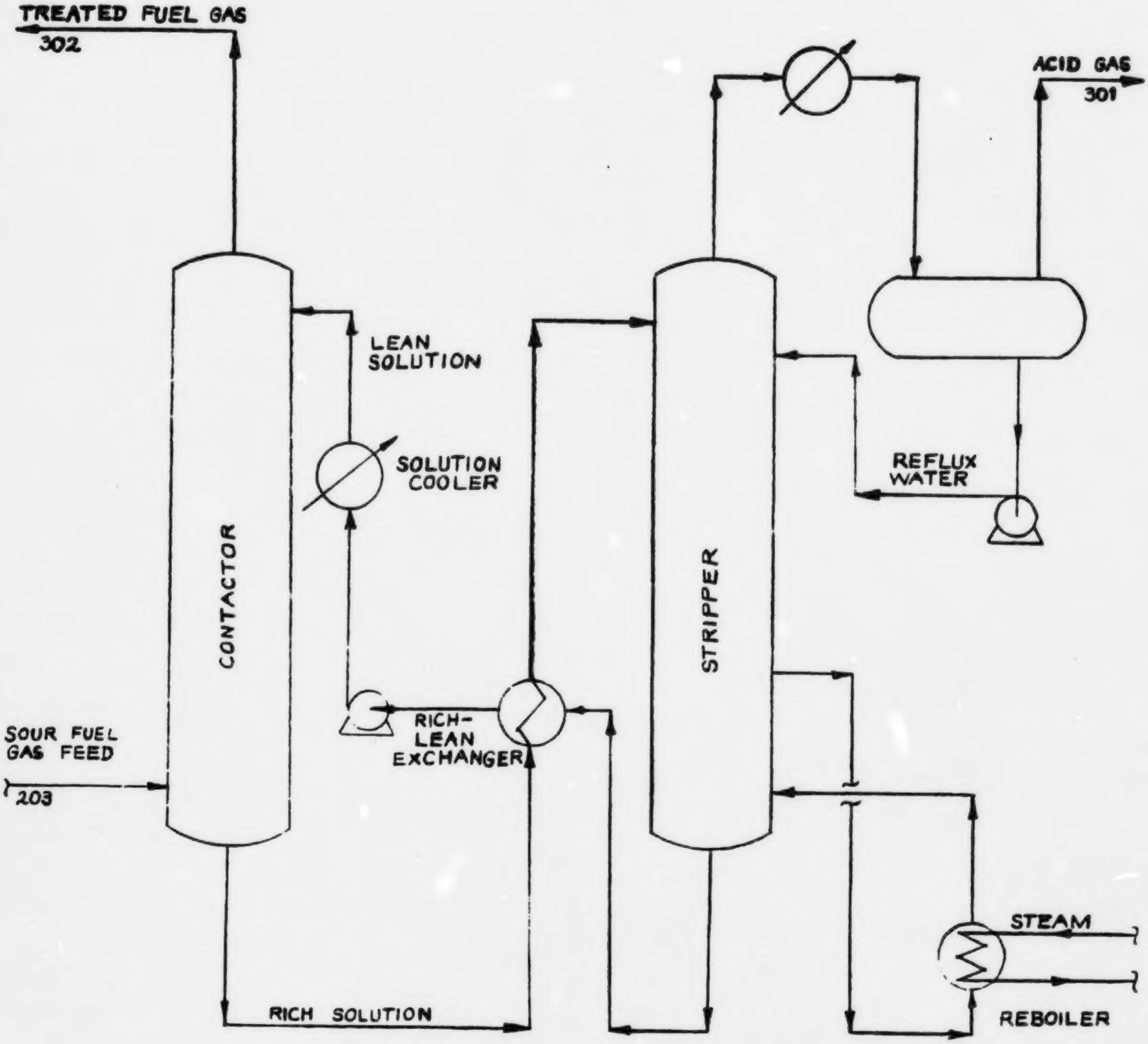

TREATED FUEL GAS
302
ACID GAS
301
LEAN
SOLUTION
SOLUTION
COOLER
REFLUX
WATER
CONTACTOR
STRIPPER
SOUR FUEL
GAS FEED
203
RICH-
LEAN
EXCHANGER
STEAM
REBOILER
RICH SOLUTION

FIGURE 4.5.1.2
CONVENTIONAL AMINE UNIT

DYNAWEST PROJECTS LTD.

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET CONT SHT

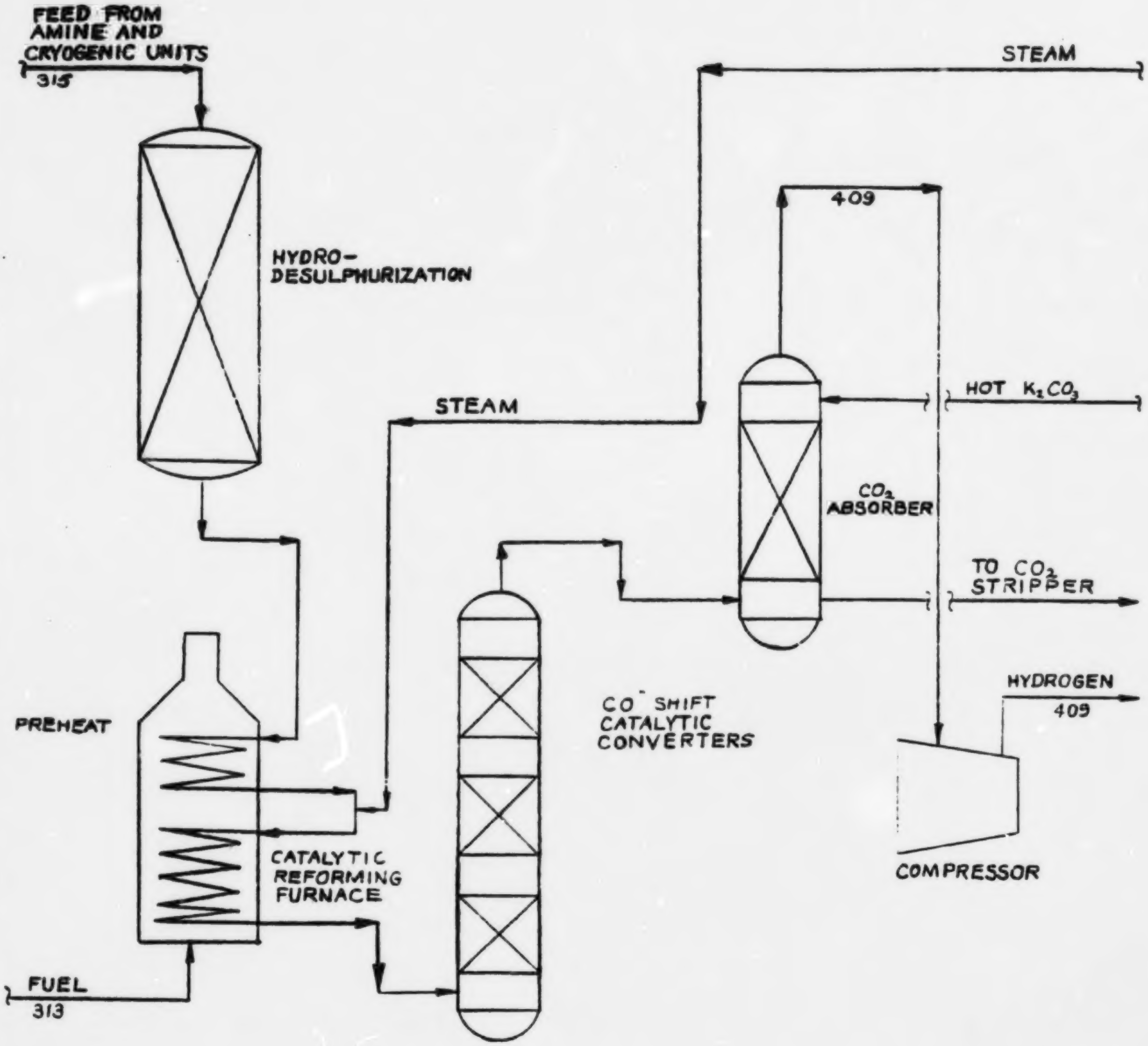

FEED FROM AMINE AND CRYOGENIC UNITS
315
HYDRO-DESULPHURIZATION
STEAM
409
STEAM
HOT K_2CO_3
CO_2 ABSORBER
TO CO_2 STRIPPER
PREHEAT
CO SHIFT CATALYTIC CONVERTERS
HYDROGEN
409
CATALYTIC REFORMING FURNACE
COMPRESSOR
FUEL
313

FIGURE 4.5.2.1
REFORMING AND SHIFT TRAIN

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

DYNAWEST PROJECTS LTD.

SHEET

CONT SHT

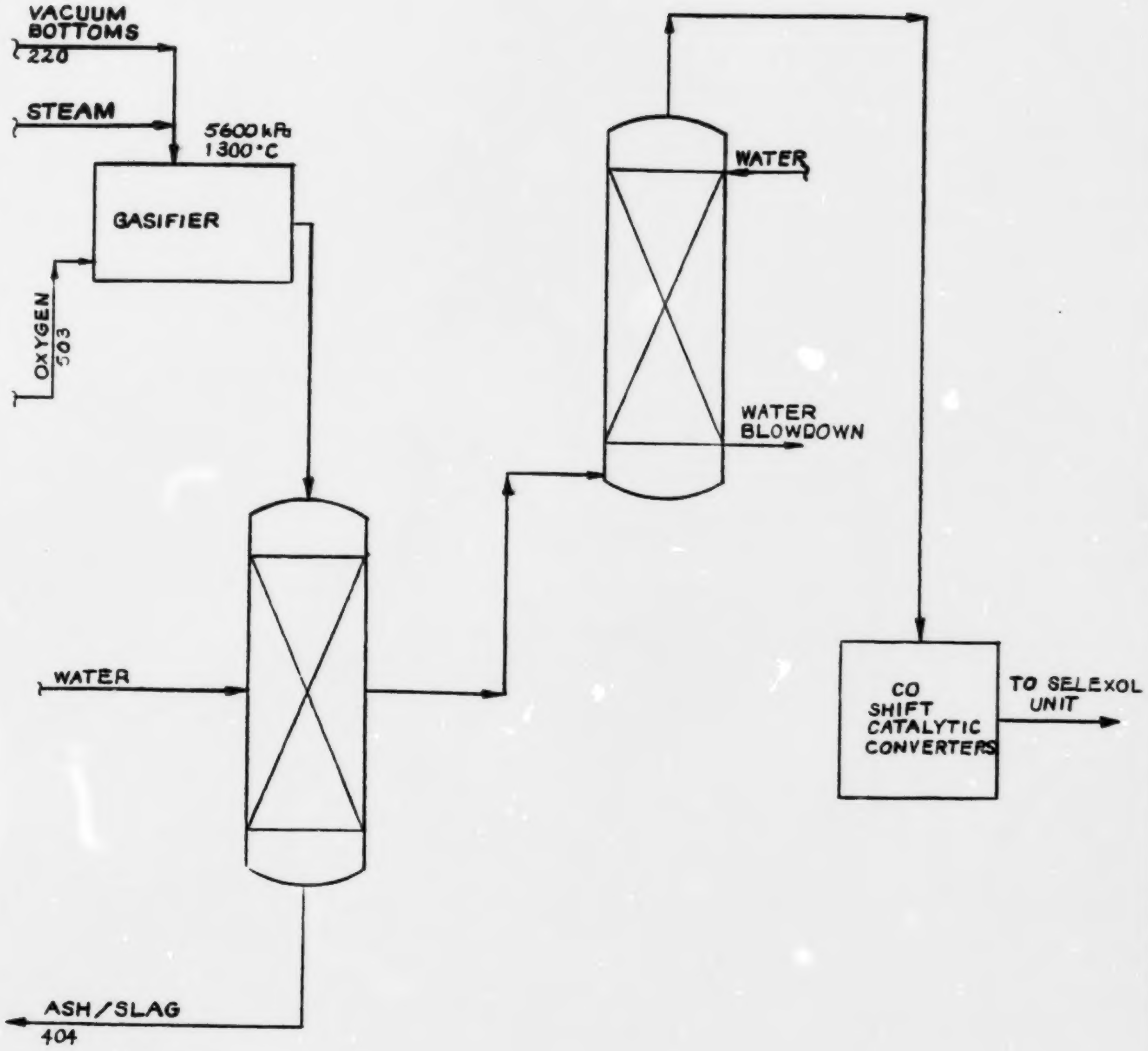

VACUUM BOTTOMS
220
STEAM
5600 kPa
1300°C
GASIFIER
OXYGEN
503
WATER
ASH/SLAG
404
WATER
WATER BLOWDOWN
CO SHIFT CATALYTIC CONVERTERS
TO SELEXOL UNIT

FIGURE 4.5.2.2
GASIFICATION TRAIN

DYNAWEST PROJECTS LTD.

ISSUE
DESCRIPTION
BY
CHK'D
APP'D
DATE
SHEET
CONT SHT

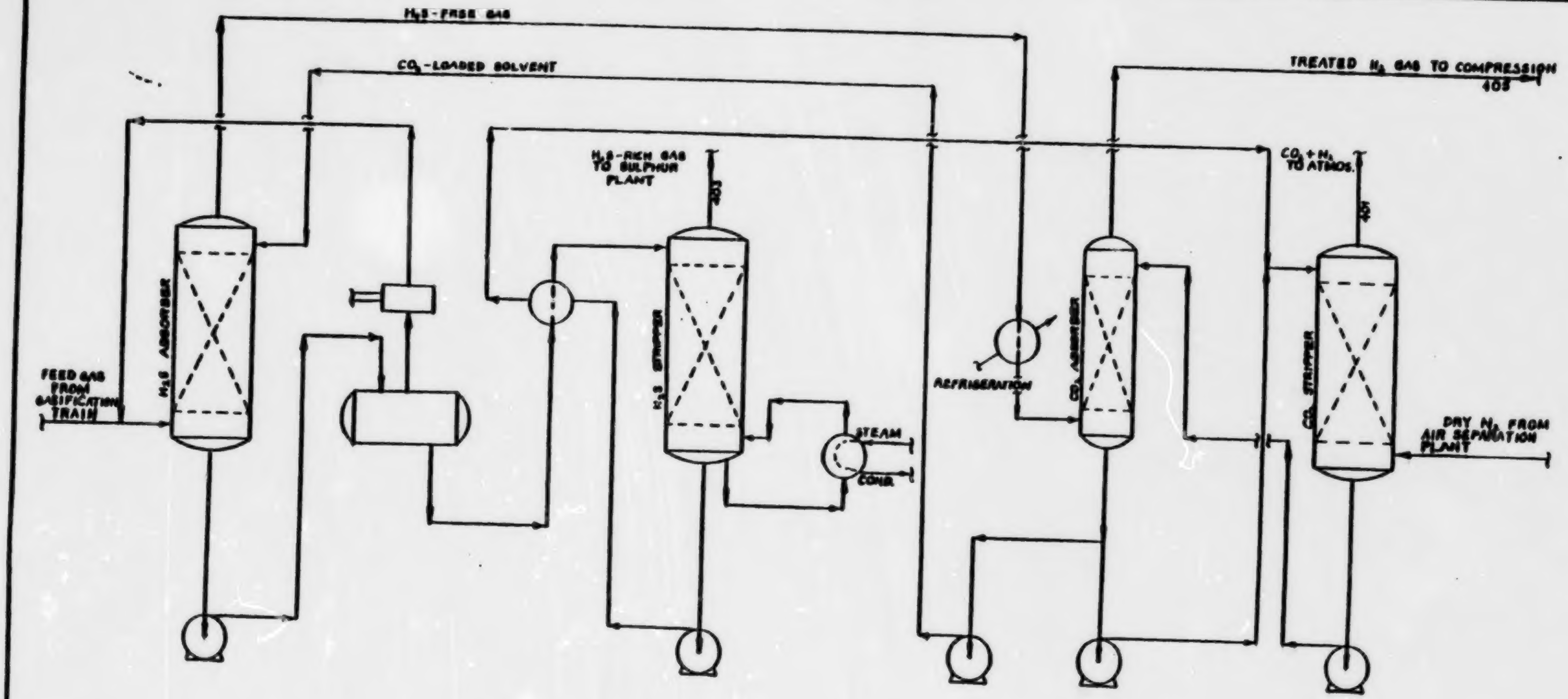

H_2S-FREE GAS
CO_2-LOADED SOLVENT
TREATED H_2 GAS TO COMPRESSION
403
H_2S-RICH GAS TO SULPHUR PLANT
402
CO_2 + N_2 TO ATMOS.
401
FEED GAS FROM GASIFICATION TRAIN
H_2S ABSORBER
H_2S STRIPPER
STEAM
CONR.
REFRIGERATION
CO_2 ABSORBER
CO_2 STRIPPER
DRY N_2 FROM AIR SEPARATION PLANT

FIGURE 4.5.2.3
SELEXOL PLANT - HYDROGEN UNIT
UNIT 400

DYNAWEST PROJECTS LTD.

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET	
						CONT. SHEET	

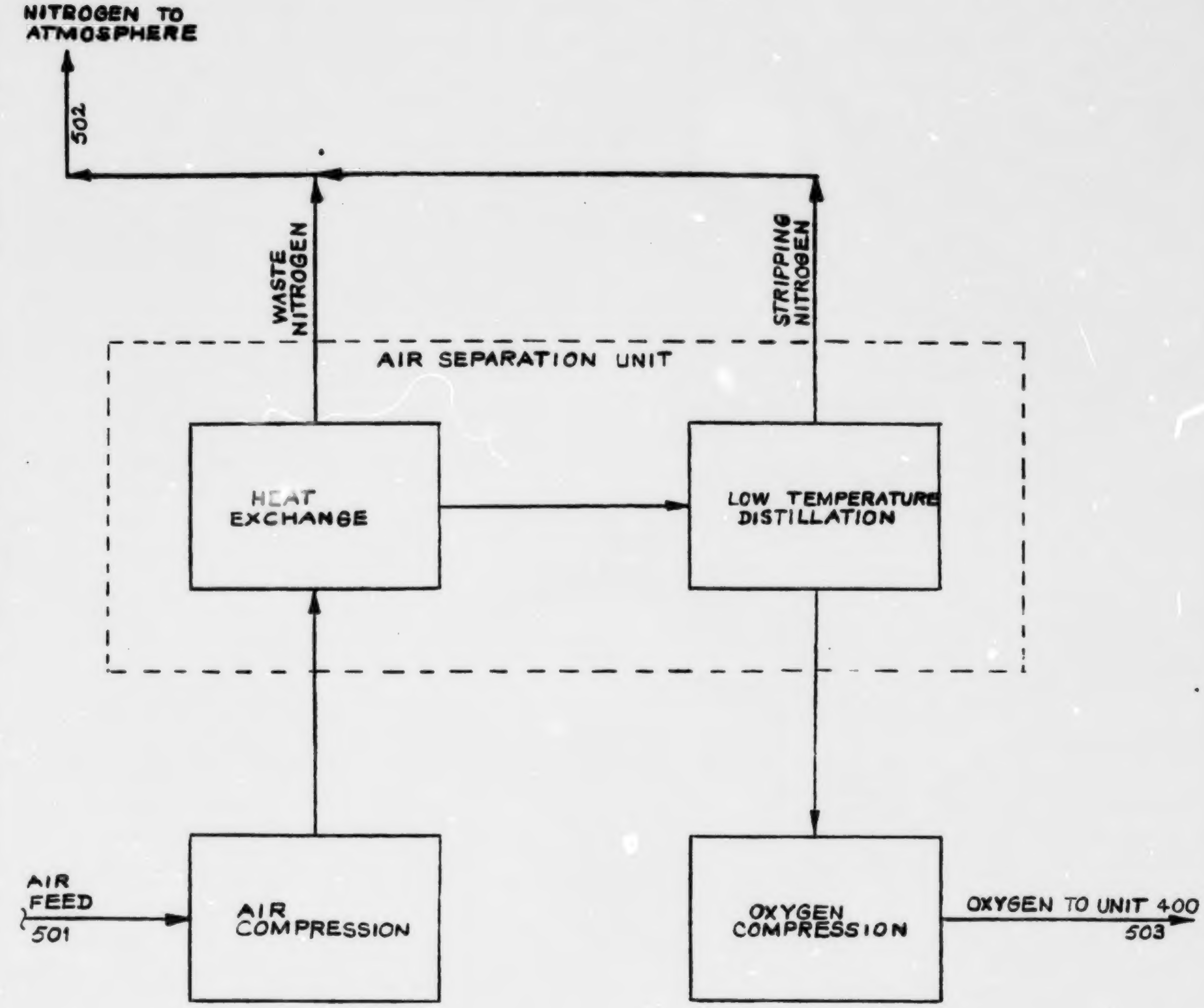

NITROGEN TO ATMOSPHERE
502
WASTE NITROGEN
STRIPPING NITROGEN
AIR SEPARATION UNIT
HEAT EXCHANGE
LOW TEMPERATURE DISTILLATION
AIR FEED
501
AIR COMPRESSION
OXYGEN COMPRESSION
OXYGEN TO UNIT 400
503

FIGURE 4.5.3
UNIT 500 – OXYGEN PLANT

DYNAWEST PROJECTS LTD.

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

SHEET
CONT SHT

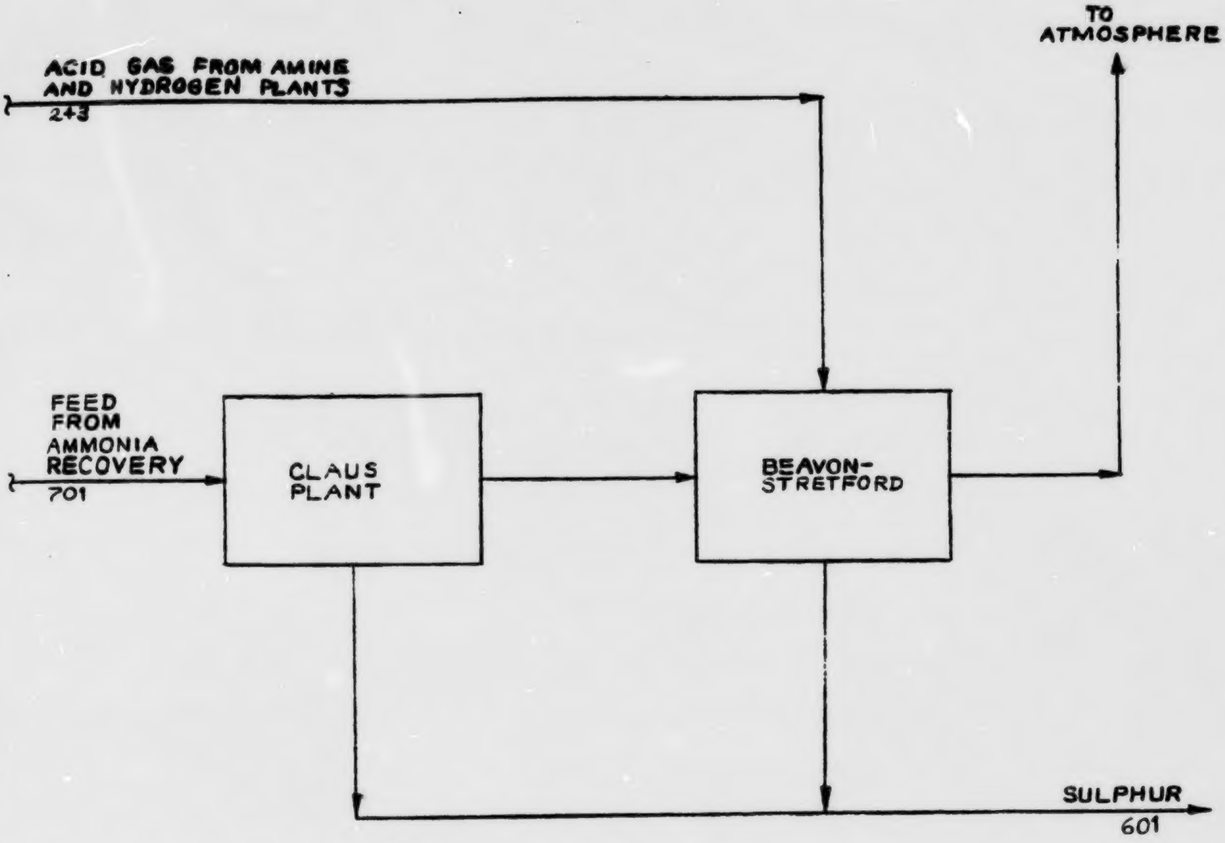

TO
ATMOSPHERE
ACID GAS FROM AMINE
AND HYDROGEN PLANTS
2+3
FEED
FROM
AMMONIA
RECOVERY
701
CLAUS
PLANT
BEAVON-
STRETFORD
SULPHUR
601

FIGURE 4.5.4.1
EMISSION CONTROL SYSTEM

DYNAWEST PROJECTS LTD.

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET CONT. SHT.

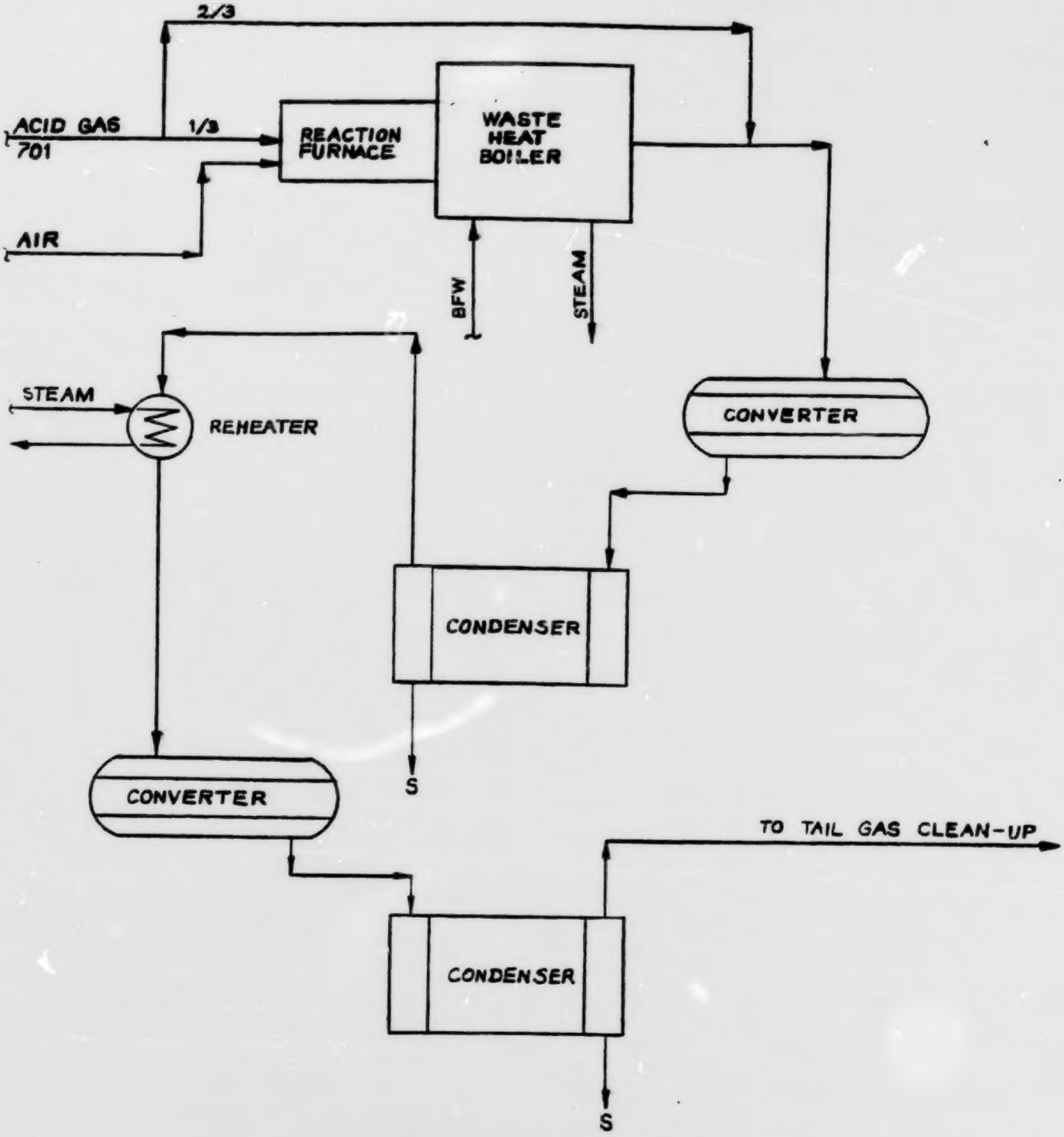

2/3
ACID GAS
701
1/3
AIR
REACTION FURNACE
WASTE HEAT BOILER
BFW
STEAM
STEAM
REHEATER
CONVERTER
CONDENSER
S
CONVERTER
TO TAIL GAS CLEAN-UP
CONDENSER
S

FIGURE 4.5.4.2
CLAUS SULPHUR PLANT

DYNAWEST PROJECTS LTD.

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE

SHEET
CONT SHT

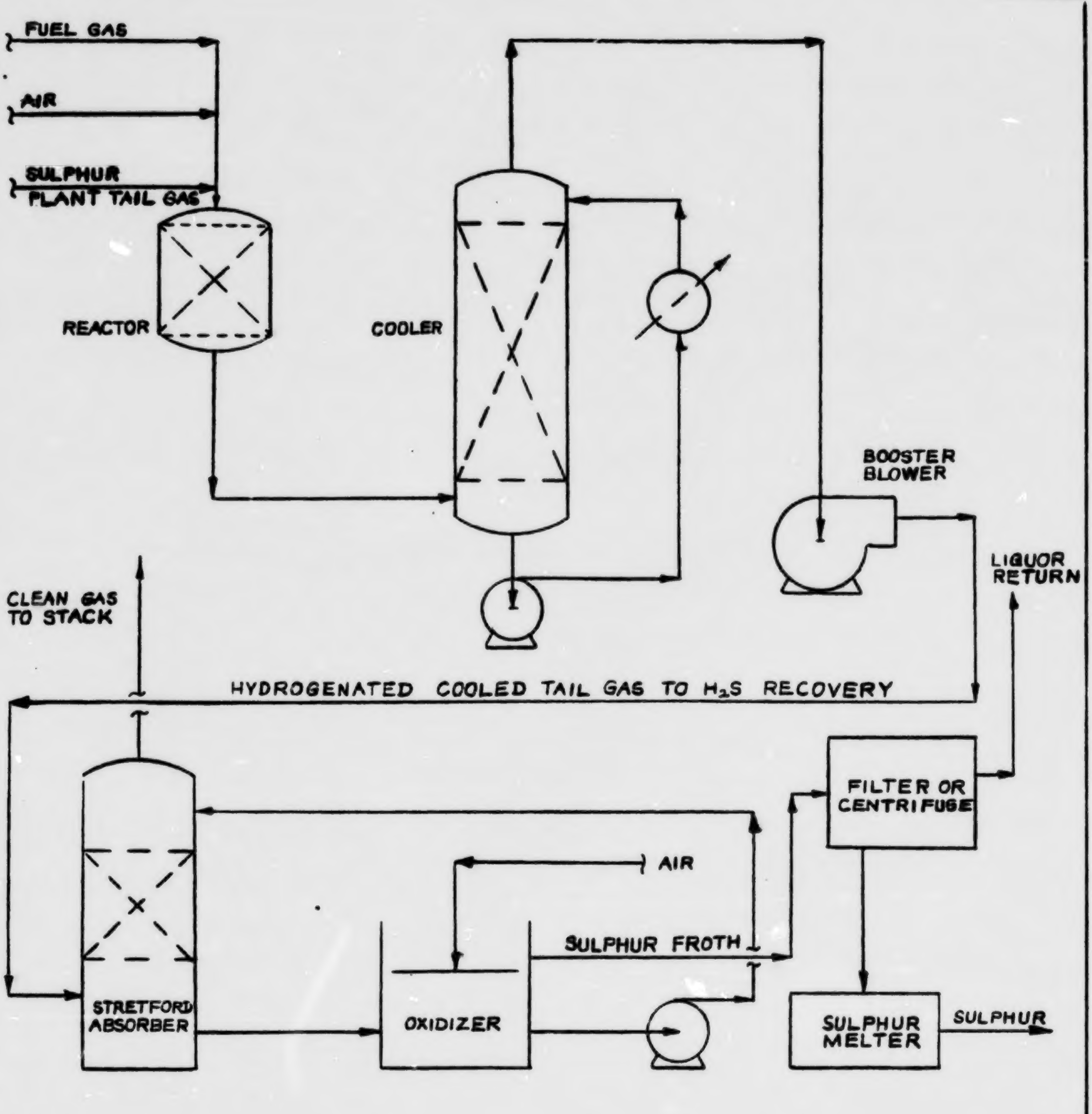

FUEL GAS
AIR
SULPHUR PLANT TAIL GAS
REACTOR
COOLER
BOOSTER BLOWER
LIQUOR RETURN
CLEAN GAS TO STACK
HYDROGENATED COOLED TAIL GAS TO H_2S RECOVERY
FILTER OR CENTRIFUGE
AIR
SULPHUR FROTH
STRETFORD ABSORBER
OXIDIZER
SULPHUR MELTER
SULPHUR

FIGURE 4.5.4.3
SULPHUR PLANT TAIL GAS TREATING
BEAVON PROCESS — UNIT 600

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	
						DYNAWEST PROJECTS LTD.
						SHEET
						CONT SHT

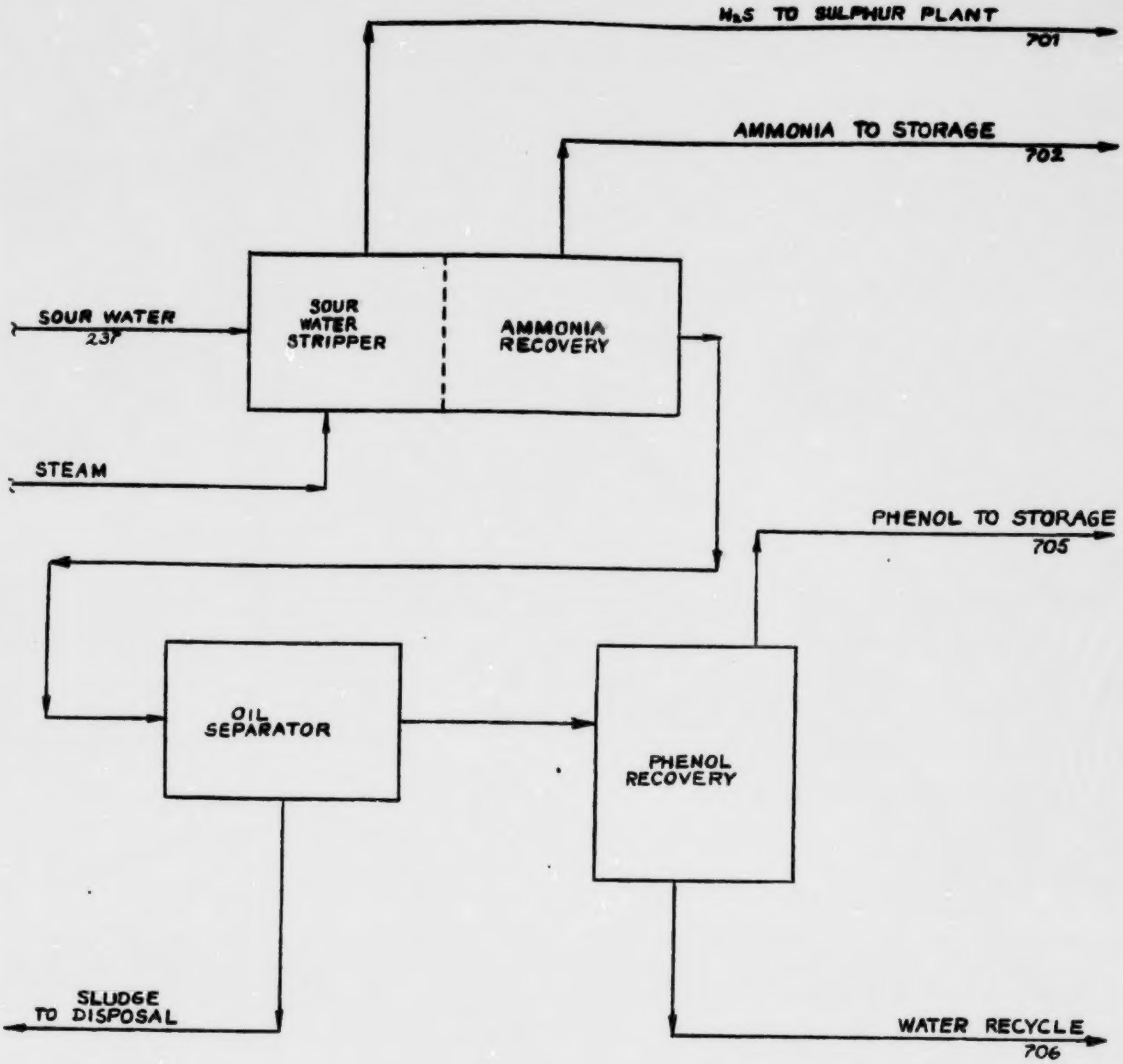

H₂S TO SULPHUR PLANT
701
AMMONIA TO STORAGE
702
SOUR WATER
237
SOUR WATER STRIPPER
AMMONIA RECOVERY
STEAM
PHENOL TO STORAGE
705
OIL SEPARATOR
PHENOL RECOVERY
SLUDGE TO DISPOSAL
WATER RECYCLE
706

FIGURE 4.5.5.1
EFFLUENT CONTROL SYSTEM

DYNAWEST PROJECTS LTD.

ISSUE	DESCRIPTION	BY	CHK'D	APP'D	DATE	SHEET
						CONT SHT

5.0 PROCESS ECONOMICS

5.1 KEY ECONOMIC SUMMARY

Within the constraints of this study and excluding infrastructure costs, we have estimated that the capital investment required for the coal based route exceeds that for the oil sands process by approximately $0.75 billion. The key reasons are illustrated in Figure 5.1 which shows relative as well as absolute values. The major difference can be seen to lie in the primary ore treatment process which in the H-Coal process comprises multi-train coal preparation and liquefaction equipment costing about twice as much as its counterpart, the extraction and primary upgrading units, in the oil sands process.

The more sophisticated coal technology requires a larger operating and maintenance staff. It also has a higher chemical demand (principally natural gas for power production and catalyst for the liquefaction reaction). These combine to give a higher annual operating cost of about $270 million.

5.2 DESCRIPTION OF PROCESS COSTS

The major products of each process are summarized in Table 5.1. The quality of the hydrocarbon fractions has been specified in Section 2.0 of this report. Capital investment is summarized in Table 5.2 and annual operating costs in Table 5.3.

Operating costs are presented for an average year and may well be higher in the initial years of operation when typical start-up problems may be expected. Costs for the oil sands case are mainly based on the Alsands ERCB application(12). Costs for the coal mine are based on vendor quotes for major items and reference(30) for ancilliary equipment. The costs for the liquefaction process are based on the Fluor EPRI reports(16).

To arrive at mid 1982 costs the excalation factors given in Table 5.4 were used. These were derived by consideration of the Nelson Refinery Construction Cost index, Chemical Engineering Plant Costs index and the Marshall and Stevens index. Labour escalation over the same period was based on the "Means" Construction Cost index.

Where appropriate, an exchange rate of 1.22 was used to convert U.S. to Canadian dollars. An additional 5% was added to all Wyoming based capital costs to reflect the need for winterization protection in Alberta.

Direct field costs for the oil sands process are identified in Table 5.5 and for the total liquefaction process in Table 5.6. Table 5.7 shows how the capital investment for the coal route is split between the mine and the process equipment.

Mining and process operating costs are separately identified in Table 5.8 in order to facilitate a comparison between the two technologies. Note that for the coal mine which may be considered as a separate corporate entity from the process revenue for the sale of coal at $15/t is included and an identical expense is added to the process costs.

The coal process has not been operated commercially and the specific coal defined for this study has not been run experimentally at any scale. It should, therefore, be expected that the cost data for the coal process is less accurate than for the oil sands process although a specific figure cannot be put on this.

A detailed account of the calculations involved in producing the estimate is provided in the appendix of this report.

Table 5.1 **Summary of Major Products (kt/a)**

	Oil Sands Process	H-Coal Process
Naphtha	1 029	1 910
Distillate/Turbine Fuel	1 126	1 810
Gas Oil	2 675	1 110
LPG	110	0
Sulphur	280	50
Ammonia	0	173
Phenols	0	19

NOTE: 300 operating days/a.

Table 5.2 **Summary of Capital Investment ($ x 10^6)**

	Oil Sands Process	H-Coal Process (including mine)
Direct Field Costs	2 767	3 331
Indirect Field Costs	1 878	1 986
Preproduction Costs	918	887
Post Start-Up Costs	180	-
Working Capital	102	198
Interest During Construction	1 500	1 669
Subtotal	7 345	8 071
Major Infrastructure Costs	963	-
TOTAL	8 308	8 071

NOTE: Indirect costs include home office costs, fees and contingency.

Sources: Oil Sands Process (12)
Coal Mine Vendor quotes and (30)
H-Coal Process (16)

Table 5.3 **Summary of Annual Operating Costs ($ x 10^6/a)**

	Oil Sands Process	H-Coal Process (including mine)
Wages and Salaries	257	324.4
Operating Materials, Chemicals	56.6	161.6
Maintenance Materials	89.7	152.9
Corporate Overhead	24.8	33.5
Property Tax and Insurance	40.0	56.2
Reclamation and Revegetation	6.0	5.5
Imported Power	3.9	15.9
TOTAL	478.0	750.0

Sources: Oil Sands Process (12)
Coal Mine
H-Coal Process (16)

Note: Costs are estimated for an average year.

Table 5.4 **Escalation Figures**

	Capital	Labour	Remarks
1978	6%	6%	From mid 1978
1979	10%	8%	
1980	10%	8%	
1981	10%	8%	
1982	6%	5%	To mid 1982
OVERALL	50%	40%	

Table 5.5 Oil Sands Process (Direct Field Costs, \$ x 10^6)

	Materials & Equipment	Labour/Subcontracts and other Construction Costs	Total
Mining	502.5	247.2	749.7
Extraction	139.3	69.1	208.4
Upgrading:			
Diluent Recovery	34.5	13.8	48.3
Primary Upgrading	183.5	73.4	256.9
Oxygen Plant	74.5	30.1	104.6
Hydrogen Plant	158.7	63.1	221.8
Hydrotreating	176.6	70.7	247.3
Gas Treating	22.1	9.4	31.5
Sulphur Plant	27.6	11.6	39.2
Utilities	178.0	71.6	249.6
Offsites and Storage	405.7	203.4	609.1
TOTAL	1 903.0	863.4	2 767.0

Source: Reference (12)

Table 5.6 H-Coal Process (Direct Field Costs $ x 10^6)

	Materials & Equipment	Labour/Subcontracts and other Construction Costs	Total
Mining	323.0	72.0	395.0
Coal Preparation	142.8	49.8	192.6
Liquefaction and Fractionation	780.2	208.4	988.6
Oxygen Plant	156.7	54.4	211.1
Hydrogen Plants	428.9	99.5	528.4
Hydrotreating	241.1	91.9	333.0
Gas Treating	18.6	5.5	24.1
Sulphur Plant	15.0	6.5	21.5
Utilities	251.4	68.9	320.3
Offsites and Storage	213.9	102.3	316.2
TOTALS	2 572.0	759.0	3 331.0

Sources: Coal Mine Vendor quotes and (30)
H-Coal Process (16)

Table 5.7 **Coal Process, Capital Investment, Mine-process Split**

	Mine	Process	Total
Direct Field Costs	395	2 935.8	3 331
Indirect Field Costs	111	1 874.8	1 986
Preproduction Costs	105	782.1	887
Working Capital	32	165.6	198
Interest During Construction	118	1 551.4	1 669
TOTALS	761	7 309.7	8 071

Sources: Coal Mine Vendor quotes and (30)
H-Coal Process (16)

Table 5.8 Annual Operating Costs, showing mine/process split

	OIL SANDS			COAL		
	Mine	Process	Total	Mine	Process	Total
Wages and Salaries	115.7	141.3	257.0	34.8	289.6	324.4
Operating Materials	7.7	19.9	27.6	7.8	45.0	52.8
Chemicals	0.0	29.0	29.0	0.0	108.8	108.8
Maintenance Materials	25.1	64.6	89.7	7.8	145.1	152.9
Corporate Overhead	6.9	17.9	24.8	4.5	29.0	33.5
Property Tax and Insurance	11.2	28.8	40.0	7.3	48.9	56.2
Reclamation and Revegetation	6.0	0.0	6.0	5.5	0.0	5.5
Imported Power	1.1	2.8	3.9	10.0	5.9	15.9
Coal Sales	0.0	0.0	0.0	(291.0)	291.0	0.0
TOTALS	173.7	304.3	478.0	(213.3)	963.3	750.0

Sources:	Oil Sands Process (12)
	Coal Mine (30)
	H-Coal Process (16)
Note:	Costs are estimated for an average year.

Table 5.9 Specifications for Economic Calculations to be Performed by CANMET

---

1. Determine the economic feasibility of each technology using two investment criteria, as follows:

   a) The Net Present Value of cash flow using 10%, 15% and 20% as discount rates.

   b) The Rate of Return (real) on equity using debt-equity ratios of 50:50 and 75:25 and assuming both independent and subsidiary corporate structures (to show the effects of flow-through arrangements for deductible expenses).

2. Use real Canadian dollars for costs and prices expressed in the same base year.

3. Use the reference product equivalent value as a basis for coal liquid prices.

4. Include sales of by-products in the revenue streams only where these are definitely marketable.

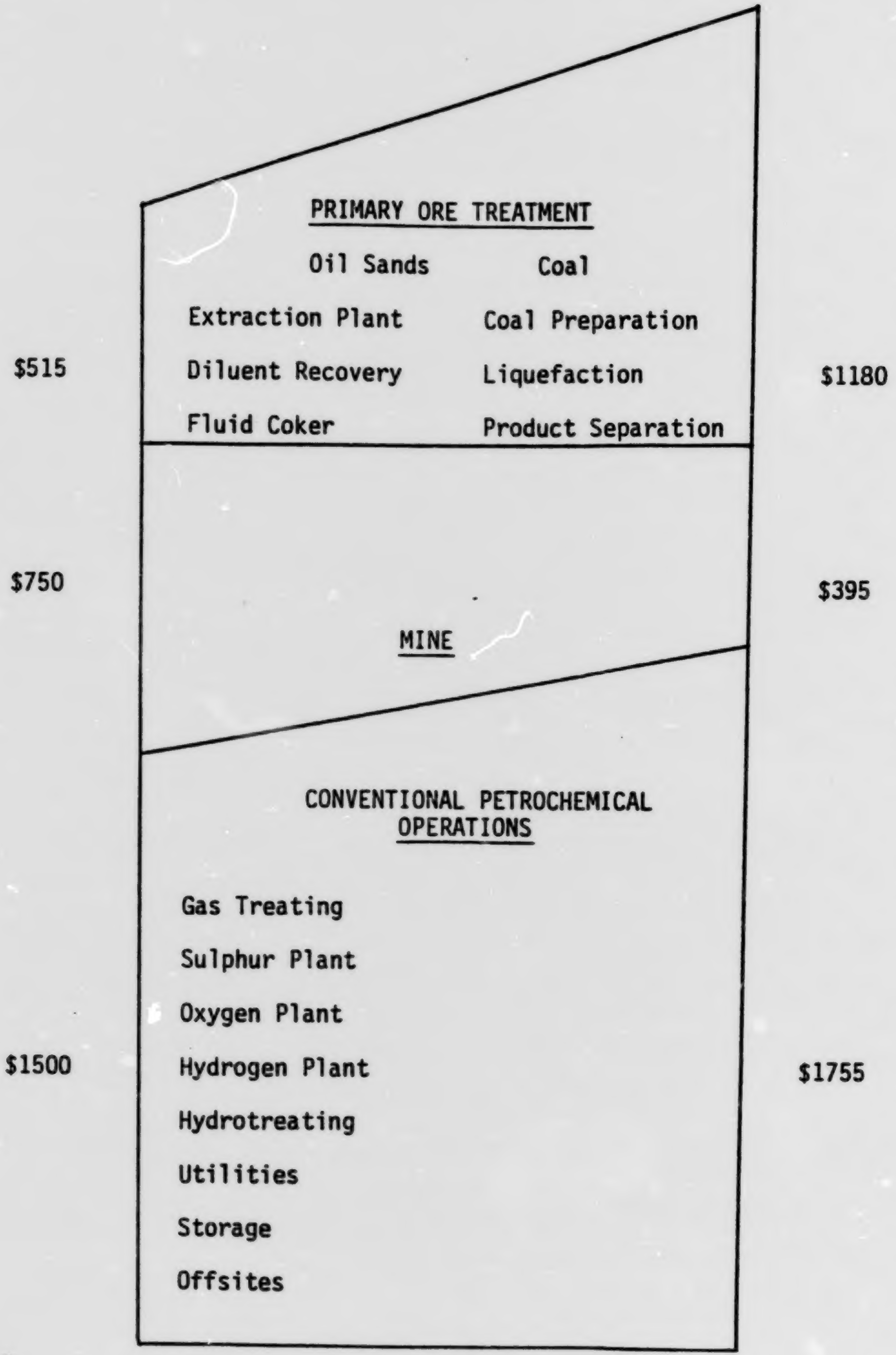
5. Costs should reflect Canadian conditions.

6. Explain the details of royalty and tax systems and any assumptions underlying the analysis.

7. Present results in table form for ease of comparison.

FIGURE 5.1

PICTORIAL REPRESENTATION OF CAPITAL COSTS


PRIMARY ORE TREATMENT
Oil Sands
Coal
Extraction Plant
Coal Preparation
$515
Diluent Recovery
Liquefaction
$1180
Fluid Coker
Product Separation
$750
$395
MINE
CONVENTIONAL PETROCHEMICAL OPERATIONS
Gas Treating
Sulphur Plant
Oxygen Plant
$1500
Hydrogen Plant
$1755
Hydrotreating
Utilities
Storage
Offsites

OIL SANDS

Direct Costs
$2765 million
Total Costs
$7345 million

COAL

Direct Costs
$3330 million
Total Costs
$8070 million

6.0 PROCESS COMPARISONS

6.1 INTRODUCTION

This section of the report compares equivalent sections of the two processes and highlights differences between the two. Alternative approaches, particularly potential improvements, to each stage are also discussed.

6.2 BASIC ORE TREATMENT

The process stages between the receipt of ore from the mine and the production of a feedstock to the conventional petrochemical operations are sufficiently different as to make any direct comparison meaningless.

Possible improvements to each process are discussed below.

6.2.1 The Oil Sands Extraction Process

The Clark Hot Water Extraction Process with its large water demand and associated tailings pond has long been regarded as the weak link in the oil sands process and the target of specific comment by the ERCB. The tailings pond is both an environmental concern and an effective method of sterilizing recoverable bitumen reserves.

The processes available for extraction of bitumen from oil sands can generally be classified into the following major groups:

1) Modified Hot Water
2) Direct Retorting
3) Solvent Extraction
4) Others

These major groups encompass all the available extraction processes currently being developed. The processes are listed in Table 6.2.1.

Reference (25) gives a general overview of the current state of technology. The RTR, Two-Stage Floatation, UMATAC Taciuk, Lurgi, Magna, SESA, Dravo and Kruyer processes are generally regarded as the most promising.

Modificiations to the hot water process itself have two main aims. The first is to improve the efficiency of the process with low grades of ore (see Figure 3.3.2). The second is to solve the tailings problem by avoiding the process conditions, in themselves not fully understood, which create the stable sludge.

The direct retorting processes involve heating the oil sand such that thermal cracking and distillation occur thus producing a vapour product and leaving a coke residue on the sand. This residue can be burnt off the sand to provide process energy. The principal advantages of retorting processes are their relative insensitivity to ore grade and the solid rather than aqueous byproduct. Heat recovery from the sand is one of the main development targets in research for these processes.

Solvent extraction processes rely on dissolving the bitumen without disturbing the water film which supposedly envelopes the sand particles. The success of any such process is dependent on the efficient recovery of the solvent.

6.2.2 Oil Sands Primary Upgrading Processes

The essential purpose of the primary upgrading stage is to reduce the viscosity of the bitumen to a level that enables it to be pumped to a refinery. This is achieved by increasing the H/C ratio. The two fundamental approaches to this are either carbon removal or hydrogen addition. The main commercial upgrading processes are listed in Table 6.2.2. All existing and proposed oil sands plants have used one of the carbon removal processes.

Delayed coking was chosen by Suncor but gives a relatively low liquid yield that would make future use unlikely. Syncrude and the proposed Alsands plant both use Fluid Coking which has been chosen as the process to be used in this study. The proposed Cold Lake project would have utilized a flexicoker.

In their detailed review of upgrading alternatives the Alsands consortium concluded that only three processes; Fluid Coking, Flexicoking and hydrovisbreaking merited detailed investigation for technical and/or economic reasons.

Although Fluid Coking was selected essentially on economic grounds it is the only process of the three that produces a significant SO_2 emission. More stringent environmental legislation will add to the cost of Fluid Coking units (in the shape of flue gas desulphurization units) and make the alternatives more competitive.

It should be mentioned that one of the two leading hydrovisbreaking processes is the H-Oil process from which the H-Coal process has been developed.

6.2.3 The H-Coal Reactor

Current HRI projections for commerical plants(23) suggest that the reactor diameter will initially be limited to about 3.5 m, i.e., a similar size to that employed on the H-Oil process from which the liquefaction process has been derived. This limitation results in a requirement of 27 parallel reaction streams for the plant capacity selected for this study.

In practice it must be expected that before a plant of this capacity is built, further development work will provide the confidence to scale up to larger diameters thus reducing the number of reactor trains required.

Among the problems predicted for the H-Coal process were mechanical problems associated with the handling of slurries at extreme temperatures and pressures and in the solid-liquid-vapour separations required downstream of the reactor. Some of these problems have now been tackled on the HRI pilot plants. HRI specifically claim to have overcome the following problems(24):

- Willis letdown valves
- EPG ball block valves
- Instrumentation - especially level controls
- Seal oil system - pump seal failure
- Isolation and maintenance valve strategy
- High pressure absorber operation
- Inadequate line tracing
- Flaker reliability
- Weigh feeder operation
- Ebullating pump vane and diffuser assembly failure.

The very nature of the process makes it unreasonable to expect that no other problems will be encountered. Equally, however, the process has now been demonstrated at a large enough scale to provide the confidence to claim that such problems as do arise will not be insuperable.

6.3 PRODUCT HYDROTREATING AND PRODUCT QUALITY

6.3.1 General

The liquid products from each plant require upgrading by hydrotreating to reduce sulphur, nitrogen and oxygen concentrations to levels that can be tolerated in refinery feedstocks.

The precise product quality required from a particular unit will be determined by the marketing and refining strategy envisaged at the design stage. In this study the products produced by current generation oil sands plants are taken as the bench mark and the products from the coal liquefaction plant are upgraded to a comparable or better quality.

Comparison of the product quality obtained from the two processes is complicated by the fact that they have not been reported on a consistent basis. The overall composition of oil sands synthetic crude is defined on Table 6.3.1. The component fractions of the coal derived synthetic crude prior to hydrotreating are defined with their properties on Table 6.3.2. These two sets of data are summarized as total synthetic crude boiling point curves on Figure 6.1. The component fractions defined in the flowsheets presented in Sections 3.0 and 4.0 are marked on this Figure.

Of particular note on Figure 6.1 is the generally lighter nature of the coal derived material. This is a consequence of the processing scheme adopted for the coal liquefaction process in which heavy material (the vacuum tower bottoms) is recycled to provide gasification feedstock for hydrogen production. Alternate processing schemes could be devised in which natural gas is used for hydrogen production, thus increasing the concentration of higher boiling materials in the synthetic crude.

Tables 6.3.3 to 6.3.6 define the properties of fractions derived from oil sands synthetic crude.

Definition of the properties of fractions prepared from Picardville coal derived liquids must perforce be more speculative, since published data on the hydrotreating of coal liquids is somewhat limited.

UOP(17) have published data for the hydrotreating of an H-Coal naphtha defined in Table 6.3.7. As can be seen, this material has a lower API gravity than the material defined by the EPRI flowsheet on which this study is based. The coal(s) from which this naphtha was prepared is not identified.

Chevron(66) have published data for the hydrotreating of an H-Coal synthetic crude containing both naphtha and distillates. Two crudes were examined. One was derived from Illinois No. 6 bituminous coal, the other from Wyodak sub-bituminous coal (which has been used as the marker for Picardville coal in this study). The latter is presented on Table 6.3.8. As would be expected, the properties of this fraction lie between those of the naphtha and turbine fuels defined by the EPRI flowsheet.

UOP(18) have published data for the hydrotreating of two H-Coal derived boiler fuels. The properties of these two fuels are defined on Table 6.3.9. They appear to bracket the properties of the material defined by the EPRI flowsheet.

Having reviewed the data available on the hydrotreating of H-Coal derived liquids, it now becomes pertinent to assess the probable properties of the fractions derived from Picardville coal.

6.3.2 Naphtha Fractions

The UOP data showing untreated and hydrotreated H-Coal naphtha is presented on Table 6.3.7. The objective of the UOP study was to demonstrate that an adequate platforming feedstock could be produced. UOP demonstrated experimentally that the hydrotreated naphtha satisfied this criteria. Extrapolating these results to the Picardville derived H-Coal naphtha, we might expect the results presented on Table 6.3.10 together with the properties of the oil sands naphtha.

The most significant property of the coal derived naphtha is higher aromatic content and consequently higher octane number. It is an excellent reformer feedstock, easily reformed to high octane levels. This is not the case for oil sands derived naphtha whose high paraffin content makes it extremely difficult to reform and necessitates blending with conventional napthas to form an acceptable feedstock to existing refineries.

While it is clear that the coal derived naphtha is a more valuable material, the extent to which this will be reflected in the market value of the material is uncertain.

6.3.3 Turbine Fuel/Distillate Fractions

In the Chevron work(66) a naphtha/turbine fuel blend was hydrotreated. The hydrotreated product was then fractionated. The properties of the 120°C+ material are presented on Table 6.3.11 together with the properties of oil sands derived kerosene and the specification for Jet A fuel. Table 6.3.11 shows that specification Jet fuel can be produced from H-Coal liquids. The same properties are presented on Table 6.3.12 together with the properties of oil sands derived diesel and the specification for No. 1D and No. 2D diesel fuel. The 120°C+ coal derived liquids met all specifications for No. 1D fuel oil. The specification for No. 2D fuel oil can be met by cutting the hydrotreated product at 150°C rather than 120°C to raise the flash point.

Inspection of Tables 6.3.11 and 6.3.12 gives the impression that coal derived liquids are superior to oil sands derived liquids for use as jet and diesel fuels. This is somewhat misleading since the data on which these tables is based was derived from experiments designed to produce a more finished product than that obtained from an oil sands plant.

6.3.4 Distillate Boiler Fuel/Gas Oil Fractions

UOP data(18) for the hydrotreating of an H-Coal derived gas oil is shown on Table 6.3.13. This data was extrapolated to the boiler fuel produced by the EPRI flowsheet. The properties of the coal and oil sands derived boiler fuels are shown on Table 6.3.14. For use as a boiler fuel, the most significant parameter is the sulphur content of the fuel which is very much lower for the coal liquids. Once again this is misleading because the UOP data on which this table is based was derived from experiments designed to produce Fluid Catalytic Cracker feedstock rather than straight boiler fuel.

6.3.5 Summary

A precise comparison of the upgrading required to convert coal liquids to products comparable to those from an oil sands plant is hampered by the lack of suitable and consistent data. Such data as is available has been derived from experimental work generally designed to produce coal liquid products of a higher specification than the products of current generation oil sands plant. The coal liquefaction flowsheet presented in this report (and in particular, the hydrogen consumption) is based on this data. This means that we have certainly overestimated the cost of producing products of comparable quality to those of an oil sands plant. This conservatism in the hydrotreating area should not have a significant effect on the overall economics as the hydrogen demand for hydrotreating is only about 17% of the total. Even a 100% over-estimate only alters total hydrogen demand by 8.5%.

The experimental data discussed above provides sufficient evidence that the H-Coal liquids can be upgraded to a quality comparable to that of the oil sands liquids. Further extensive experimental and design work would be required to clearly define (let alone optimize) the product yield from a coal liquefaction plant based on Picardville coal.

6.4 HYDROGEN PRODUCTION

The coal process has a hydrogen demand ten times that of the oil sands process and this dictates a more dominant role for the hydrogen plants in the former case. The oil sands process which incorporates a carbon removal process for the primary upgrading of the bitumen only requires hydrogen for the hydrotreaters. This hydrogen can be conveniently produced by the gasification of fluid coke. The amount of coke produced is close to precisely the amount required for hydrogen production. A detailed study would be required to confirm that the coke supply is in fact adequate for the final upgrading scheme selected. A coke shortfall could be made up by either incorporating supplementary hydrogen production from natural gas or importing surplus coke from another oil sands process. (The Syncrude plant now has a coke stockpile sufficient to supply several years production).

Although gasification of coal is an established technology, gasification of oil sands fluid coke has yet to be demonstrated at a commercial scale. The alternative method of hydrogen production, used by Syncrude and Suncor, is the steam reforming of natural gas. This results in a lower overall process utilization of raw materials since the coke not utilized is most likely to be stockpiled (as at Syncrude).

The coal process requires hydrogen for increasing the basic H/C ratio of the coal as well as for product upgrading. The total hydrogen demand is ten times that of the oil sands process. This cannot be met by reactor residue gasification alone and some form of supplemental production (e.g., steam reforming of plant fuel) is required. While the alternative to residue gasification is once again steam reforming of natural gas, the relatively large hydrogen demand makes the coal process far more dependent on the incorporation of gasification technology to achieve a reasonable overall raw material efficiency. The large demand makes hydrogen production relatively insensitive to the hydrotreating demand which only accounts for 17% of the total.

As with the reaction section current technology dictates that the gasification section of the coal process will incorporate a large number of parallel trains (nine).

6.5 GAS TREATMENT

Both processes use conventional amine technology to sweeten light hydrocarbon gases. The coal process also requires a cryogenic unit to separate hydrogen for recycle to the reactor.

6.6 SULPHUR RECOVERY

Both processes rely on a Claus plant to recover the bulk of the sulphur. Tail gas treatment strategies differ. The oil sands process simply employs a SCOT unit treating Claus plant tail gas. The coal process which has a greater diversity of H_2S rich process streams including several with concentrations below the viable level for a Claus plant employs a more elaborate tail gas treatment scheme with some of the plant sour gases being introduced directly to the Stretford unit.

6.7 EFFLUENT TREATMENT

Both processes employ sour water stripping sections of similar capacity. The higher production rates of ammonia and phenols in the coal case make the incorporation of additional units to separate these byproducts economic.

6.8 UTILITIES

6.8.1 Electrical Power

Electrical power supply for the oil sands plant is supplied by the utilities section. Power is generated from high pressure steam from CO boiler and heat recovery boilers and directly from gas turbines. There is sufficient electrical power produced for all areas including the mine. The only import power required is for start up or partial upgrading shutdown and may be obtained from the provincial power grid or by importing natural gas to run the power plant.

For the coal plant, electrical power produced in the utilities section for the coal liquefaction facility uses fuel gas and natural gas in a combined cycle power plant. Additional power is imported from the provincial grid for the mine area.

It would be possible to import additional natural gas to raise power for the mine. This seems an unreasonable strategy since the plant is located close to the main provincial electricity generation utilities whose economies of scale must logically provide the preferred source of additional power.

6.8.2 Raw Water Supply, Storage and Treatment

Minimum raw water supply to the oil sands plant is about 70,000 m^3/d with full recycle in the water management system. When recycle water from the tailings pond is unavailable, the raw water intake has to be increased to three times the normal flow rate.

The coal plant only requires a maximum of 75,000 m^3/d during the summur months. This water is mainly used for cooling tower system makeup. If air coolers and air cooled chillers are used in place of cooling towers, the water consumption could be reduced to about 15,000 m^3/d. It should be noted that a reduction in the cooling water demand for the Texaco Nanticoke refinery of this order was achieved by following this approach(31).

6.8.3 Steam Generation

For the coal plant steam is produced in the process area and combined with steam from the heat recovery steam generator. Part of the steam is consumed for electrical power generation. Multi-level steam is supplied for process use by the utilities power plant.

High pressure steam in the oil sands plant is normally raised by the CO boiler.

The oil sands process, unlike the coal process, has a large demand for low grade heat. This is met by the hot water system which recycles energy from the upgrading plant to the extraction plant rather than by a conventional steam system.

6.9 THE ENVIRONMENTAL IMPACT

6.9.1 Surface Disturbances

The two projects would cover comparable land areas approximately 100 km^2 for the coal process and 70 km^2 for the oil sands process split as shown in Table 6.9.1. However, although the coal process, particularly the mine, covers a larger area the interval between initial mining and final reclamation is shorter. More importantly at the conclusion of the project the coal mine can be completely revegetated. At the moment there is no known method for the reclamation of the tailings pond that will remain at the conclusion of the oil sands project.

6.9.2 Water Management

Both processes have a nominal water demand of 75,000 m^3/d. For the oil sands process this is the minimum rate when tailings pond recycle water is available. For the coal plant this is the maximum rate when cooling tower losses are highest and could in any event be reduced by adoption of additional direct and indirect air cooling capacity.

While the coal process has a potentially lower water demand it should be remembered that, unlike the oil sands project, it is not located close to a major water supply. In a more detailed study, the water demand of the process would have to be carefully matched against the known supply sources.

6.9.3 Atmospheric Emissions

Both processes have the potential for producing significant atmospheric emissions. Sulphur emissions generally give rise to most concern. These are much higher with the oil sands process for two reasons:

1. The sulphur content of the feedstock is higher (0.049 mass fraction c.f. 0.005 mass fraction).
2. The incorporation of the fluid coker in the upgrading process produces an SO_2 stream that on current generation oil sands plants is emitted without flue gas desulphurization.

The latter problem could be avoided in future processes by either the incorporation of a FGD unit or the adoption of a different upgrading configuration. However, the higher initial sulphur content of the ore dictates that with the inclusion of best practicable technology in each case the oil sands process will always generate higher sulphur emissions.

Both processes produce significant water vapour emissions (the oil sands process from the tailings pond, the coal process from the drying operation). In each case these give the potential for ice fog formation in winter and in each case the main remedy is to site the emission well away from public areas, particularly roads. Control of the more localized emission from the coal driers is, however, far more feasible than conrol of the area emission from a tailings pond.

Particulate emissions can arise from both processes. The main sources are the CO boiler flue gas in the oil sands process and the coal mine, handling and drying areas in the coal process. While technology is available to control particulate and dust emissions in each case it must be anticipated that these emissions will be greater from the coal based process.

6.9.4 Infrastructure and Social Impact

The location of the Athabasca oil sands deposits dictates that each new project requires a substantial investment in new service supplies and general infrastructure (e.g., power supplies, transportation system, accommodation, etc.). The coal project, however, situated close to Edmonton will not require much of this expenditure.

For the same reason the social impact of each new oil sands project must be more carefully considered than a comparable investment in the Edmonton region.

Some of the occasionally spasmodic and troubled operation of the existing oil sands plants has been attributed to the shortage of skilled operators in the Fort McMurray area and to the high turnover rate of staff in general. These problems would be expected to be less acute in the Edmonton catchment area.

6.10 PRODUCT UTILIZATION

6.10.1 General

Both the oil sands product and the coal product differ sufficiently from conventional crude oil to necessitate modification of existing refinery practice. To date the limited total production of oil sands synthetic crude has permitted their accommodation in existing refineries by blending with conventional feedstock. If the production of sythetic crude is to increase then a more positive strategy is clearly required. The first refinery specifically designed to process synthetic crude is being constructed at Scotford, Alberta by Shell Canada and is scheduled for start up in 1984(28).

Once the practical limits of blending have been exceeded two alternative strategies, or a combination thereof exist, viz:

1. Extend the synthetic crude upgrading operation to match existing refinery capability.
2. Modify existing refineries (or build new ones) to match synthetic crude properties.

The optimum strategy can only be determined in relation to the predicted synthetic crude production rate and the market demand for particular petroleum products.

6.10.2 Oil Sands Production Utilization

By comparison with conventional crude, oil sands synthetic crude characteristics of signifigance to a refinery are(26):

1. Very low sulphur content
2. Lower naphtha content
3. Higher distillate content
4. Higher gas oil content
5. Lack of short residue
6. Poorer distillate combustion quality as indicated by Jet A-1 smoke point and Diesel cetane number.

In a purpose built refinery the synthetic crude can be made to meet all current specifications for gasoline, diesel and fuel oils. References (26) and (27) contain detailed descriptions of the many options available for the incorporation of oil sands synthetic crude in existing refineries.

6.10.3 Coal Product Utilization

In broad terms, the synthetic crudes produced by coal liquefaction show the same major differences from conventional crude oil that have been identified above for oil sands synthetic crude. The most probable product disposition is identified in Table 6.10.1.

Detailed studies (17) and (18) have been published on the upgrading of specific H-Coal synthetic crude fractions. In each case intermediate or final products within normal refinery specifications were obtained by the application of conventional processing techniques. (Hydrotreating, reforming, fluid catalytic cracking). In particular, it was concluded that fluid catalytic cracking, in conjunction with hydrotreating, has the potential for playing a major role in refining the heavier coal derived synthetic crude fractions.

A further study (29) has been published detailing the most likely refining strategy for a dedicated coal liquids refinery and including an assessment of the upgrading costs incurred in converting raw coal liquids to finished refinery products.

6.10.4 Byproducts

The major byproducts of the oil sands process are LPG and sulphur. While the LPG is a marketable product it will be difficult to profitably dispose of the large sulphur production (280,000 t/a) particularly in view of the remote location of the plant.

The major byproducts of the coal process are ammonia in substantial quantities (173,000 t/a) and a small quantity of phenols. The ammonia should be relatively easy to market as fertilizer feedstock in Alberta. The phenols rate of 19,000 t/a is fairly low. If no market were anticipated the process configuration could be altered to pass the phenols to an in plant incinerator and recover the energy content thus displacing some of the natural gas import.

Although not listed as byproducts in Table 5.1, it should be remembered that the hydrogen plants in each process produce a CO_2 vent stream. Some interest has been expressed in the recovery of CO_2 from power stations as an in situ flooding medium for bitumen extraction. The coal plant with its much larger hydrogen demand obviously has a larger associated CO_2 vent.

Table 6.2.1 Alternatives to the Clark Hot Water Extraction Process

Process	Reference
Modified Hot Water Processes	
RTR	32
Two-Stage Flotation	33
Petreco	-
Sohio	34
ARC	35
Direct Retorting	
UMATAC/Taciuk	36
Mobil	37
Lurgi	38
University of Calgary (Fluidized Bed Process)	39
Solvent Extraction Processes	
Magna	-
Mineral Resources Corportation	40
SESA	41
Natomas	42
Dravo	-
Exxon	43
Guardian (DIKOR)	44
DECO	45
Gulf	46
Mobil	47
Phillips	48
Texaco	49
Super Critical Solvent	50
Technical Economists Ltd.	51
Chevron	52
Others	
Kruyer Oleophilic	53
Worne Biochemical Sieve	-
Resource Recovery Engineering	-
Ultra Sonics (Western Oil Sands Ltd.)	54

Table 6.2.2 Processes for the Primary Upgrading of Bitumen

Process	Reference
Carbon Removal Processes	
Delayed Coking	55
Fluid Coking	56
Flexicoking	56
Eureka	57
Solvent Deasphalting	58
Catalytic Cracking	59
Hydrogen Addition Processes	
Hydrovisbreaking	60
Fixed Bed Hydrodesulphurization	61
Aurabon	62
Asphaltene Cracking	63
Dynacracking	64

Table 6.3.1 Synthetic Crude from Oil Sands

Crude Quality	
API	33.0
Sulphur (mass fraction)	2×10^{-3}
Nitrogen (mass fraction)	0.7×10^{-3}
Composition (Volume Fraction)	
C_3	0.01
C_4	0.038
C_5 - 150°C	0.175
150 - 335°C	0.375
335°C+	0.402
	1.000

Table 6.3.2 Raw Synthetic Crude from H-Coal Process

	Naphtha	Turbine Fuel	Distillate Boiler Fuel
API	55.8	27.8	10.9
Sulphur (mass fraction)	0.2×10^{-3}	0.3×10^{-3}	0.3×10^{-3}
Nitrogen (mass fraction)	1×10^{-3}	2.2×10^{-3}	4.6×10^{-3}
Mass fraction of Total Synthetic Crude	0.4	0.37	0.23
Composition (Mass Fraction)			
C_4	0.01	-	-
C_5 - 40°C	0.07	-	-
40 - 95°C	0.23	-	-
95 - 150°C	0.45	-	-
150 - 205°C	0.24	-	-
175 - 315°C	-	1.00	-
260 - 425°C	-	-	1.00
	1.00	1.00	1.00

Table 6.3.3 Properties of Oil Sands Synthetic Crude Naphtha(27)

Boiling range	C_5 - 150°C
Density, kg/m^3	697
Sulphur, mass fraction	1×10^{-6}
GC Hydrocarbon analysis, volume fraction	
Paraffins	0.69
Naphthenes	0.25
Aromatics	0.06
Benzene	0.011
Toluene	0.027
Xylene	0.02
Research Octane No.	58
Motor Octane No.	57

Table 6.3.4 Properties of Oil Sands Synthetic Crude Kerosene(27)

	Synthetic	Jet A Specifications
Density, kg/m^3	830	775-839
Viscosity, cSt at 40°C	1.3	-
Freeze pt., °C	-60	-40 max
Sulphur, mass fraction	38×10^{-6}	2×10^{-3} max
Nitrogen, mass fraction	4×10^{-6}	-
Hydrogen, mass fraction (D 3701)*	0.129	-
Aromatics, volume fraction (D 1319)*	0.32	0.20 max
Naphthalenes, mass fraction (D 1840)*	0.006	0.03 max
Smoke point (D 1322)* mm	13	20 min
Luminometer no. (D 1740)*	29.3	45 min

* ASTM Test methods

Table 6.3.5 Properties of Oil Sands Synthetic Crude Diesel(27)

Density, kg/m^3	897.4
Viscosity, cSt at 40°C	3.3
Cloud point, °C	-27
Pour point, °C	-50
Sulphur, mass fraction	300×10^{-6}
Nitrogen, mass fraction	25×10^{-6}
Aromatics, volume fraction (D 1319)	0.44
Cetane no. (measured) (D 613)	34
Cetane index (calculated) (D976)	42
Distillation (D 86)	
IBP°C	157
10%	192
50%	274
90%	329
FBP	334

Table 6.3.6 Properties of Oil Sands Synthetic Crude Heavy Fuel Oil(27)

Boiling range, °C	455°C+
Yield on crude, volume fraction	0.13
Density, kg/m^3	936
Viscosity, cSt at 100°C	18
Sulphur, mass fraction	2×10^{-3}
Nitrogen, mass fraction	1.2×10^{-3}
Pour Point, °C	9
Conradson carbon residue, mass fraction	2×10^{-3}
BMCI	49

Table 6.3.7 Experimental Data for Hydrotreating H-Coal Derived Naphtha(17)

	Raw Naphtha	Hydrotreated Naphtha
API	43.7	46.8
Density kg/m^3	807.6	793.6
Distillation D-86		
IBP^{o}C	55	67
5%	77	85
10%	87	93
20%	102	103
30%	112	111
40%	122	119
50%	127	128
60%	144	140
70%	156	152
80%	164	165
90%	122	178
95%	129	186
EP	202	201
% over	99	99
Hydrogen (mass fraction)	0.128	0.1359
Carbon (mass fraction)	0.859	0.8645
Sulphur (mass fraction)	1.29×10^{-3}	3.9×10^{-6}
Nitrogen (mass fraction)	1.93×10^{-3}	0.63×10^{-6}
Oxygen (mass fraction)	5.94×10^{-3}	34.0×10^{-6}
Chloride (mass fraction)	23.0×10^{-6}	4.0×10^{-6}
FIA (Volume fraction)		
A	0.247	0.189
O	0.052	0.0
P & N	0.701	0.811
MS Hydrocarbon Types (Volume fraction)		
A	0.186	0.176
N	0.555	0.634
P	0.162	0.190
Bromine Index	19.3[a]	296
RON, Clear	80.3	66.8

[a] Bromine number.

Table 6.3.8 **Properties of H-Coal Derived Syncrude used in Hydrotreating Studies(66)**

API	35.1
Aniline point °C	1.0
Molecular weight	130.0
Viscosity cst at 38°C	1.225
Ramsbottom Carbon (mass fraction)	2.3×10^{-3}
Hot Heptane insoluble (mass fraction)	0.68×10^{-3}
Ash (mass fraction)	20.0×10^{-6}
Bromine number	26.0
Carbon (mass fraction)	0.862
Hydrogen (mass fraction)	0.127
Sulphur (mass fraction)	0.41×10^{-3}
Oxygen (mass fraction)	8.5×10^{-3}
Nitrogen (mass fraction)	1.7×10^{-3}
Iron (mass fraction)	9.0×10^{-6}
Chloride (mass fraction)	3.0×10^{-6}
H/C Atomic ratio	1.76
Water (mass fraction)	0.5×10^{-3}
TBP Distillation °C	
st 15	12/69
10/30	78/127
50	179
70/90	221/279
95/99	317/418

NOTE: This H-Coal syncrude was prepared from Wyodak sub-bituminous coal.

Table 6.3.9 H-Coal Derived Gas Oils used in Hydrotreating Studies(18)

	Sample 1	Sample 2
API	12.9	8.9
Density kg/m^3	979.9	1 008
Distillation D-1160		
IBP°C	214	213
5%	242	339
10%	249	254
20%	259	270
30%	269	281
40%	280	289
50%	292	299
60%	303	310
70%	317	324
80%	337	340
90%	365	365
95%	386	379
EP	410	409
% over	99.0	99.0
% bottoms	1.0	1.0
Hydrogen (mass fraction)	0.096	0.091
Carbon (mass fraction)	0.893	0.890
Sulphur (mass fraction)	0.5×10^{-3}	0.8×10^{-3}
Nitrogen (mass fraction)	4.2×10^{-3}	4.1×10^{-3}
Oxygen (mass fraction)	5.2×10^{-3}	5.2×10^{-3}
Conradson Carbon (mass fraction)	0.1×10^{-3}	0.1×10^{-3}
Heptane Insoluble (mass fraction)	1.1×10^{-3}	$.5 \times 10^{-3}$
FIA (Volume fraction)		
A	0.84	0.902
O	0.0	0.0
P & N	0.16	0.098

Table 6.3.10 Properties of Hydrotreated Naphthas from Oil Sands and Coal Liquefaction Plants

	Oil Sands Naphtha	H-Coal Naphtha
API	71.5	58.0
Density kg/m^3	697.0	747.0
Sulphur (mass fraction)	1.0×10^{-6}	4.0×10^{-6}
Nitrogen (mass fraction)	-	1.0×10^{-6}
Aromatics (volume fraction)	0.06	0.19
RON	58.0	67.0
MON	57.0	-

Table 6.3.11 Properties of Hydrotreated Distillate/Turbine Fuel from Oil Sands and Coal Liquefaction Plants (vs. Jet A Spec.)(27,66)

	Oil Sands Derived	Coal Derived	Jet A Specification
API	39.0	39.7	37 to 51
Density, kg/m^3	830.0	827.0	775 to 839
Freeze point °C	-60	-40	-40
Sulphur (mass fraction)	38.0×10^{-6}	20.0×10^{-6}	2.0×10^{-3} max.
Nitrogen (mass fraction)	4.0×10^{-6}		
Aromatics (volume fraction)	0.32	0.036	0.2
Naphthalenes (mass fraction)	0.006		0.03
Smoke point mm	13.0	21.0	20.0

NOTE: Coal derived properties based on material prepared from Wyodak sub-bituminous coal.

Table 6.3.12 Properties of Hydrotreated Distillate/Turbine Fuel from Oil Sands and Coal Liquefaction Plants (vs. Diesel Specifications)(27,66)

			Diesel Specification		
			No. 1	No. 2 D	
	Oil Sands Derived	Coal Derived	All Climates	Typical Climates	Cold Climates
API	26.2	39.7	-	-	-
Density, kg/m^3	897.4	827.0	-	-	-
Cloud point ^oC	-27	-29	Varies	Varies	-12
Sulphur (mass fraction)	300×10^{-6}	20×10^{-6}	5×10^{-3}	5×10^{-3}	5×10^{-3}
Aromatics (volume fraction)	0.44	0.036	-	-	-
Cetane No.	34.0	41.0	40.0	40.0	40.0
D86, 90% ^oC	329.0	259.0	288.0	282 - 293	293.0

NOTE: Coal derived properties based on material prepared from Wyodak subbituminous coal.

Table 6.3.13 Experimental Data for Hydrotreating H-Coal Derived Gas Oil(18)

	Raw Gas Oil	Hydrotreated Gas Oil
API	8.9	20.4
Density, kg/m^3	1008.0	932.0
Hydrogen (mass fraction)	0.091	0.116
Carbon (mass fraction)	0.89	0.885
Sulphur (mass fraction)	0.8×10^{-3}	0.6×10^{-6}
Nitrogen (mass fraction)	4.1×10^{-3}	95.0×10^{-6}
Oxygen (mass fraction)	5.2×10^{-3}	1.2×10^{-3}
FIA (volume fraction)		
A	0.902	0.512
O	0	0
P & N	0.098	0.488

Table 6.3.14 Properties of Hydrotreated Gas Oil from Oil Sands and Coal Liquefaction Plants(27, 18)

	Oil Sands Derived	Coal Derived
API	19.7	20.4
Density, kg/m^3	936	932
Sulphur (mass fraction)	2×10^{-3}	0.6×10^{-6}
Nitrogen (mass fraction)	1.2×10^{-3}	95×10^{-6}
BMCI	49	65

Table 6.9.1 Land Area (km^2) Required for Synthetic Fuel Projects

	Oil Sands Process	Coal Liquefaction Process
Mine	27	75
Tailings Pond	21	-
Plant	22	25
Total	70	100

Table 6.10.1 Probable Disposition of H-Coal Products (24)

Synthetic Crude Fraction	Upgrading Required	Disposition	Comments
Naphtha	Hydrotreating Catalytic reforming	Gasoline, Aromatics production	Severe hydrotreating. Excellent reformer feedstock.
Turbine Fuel	Mild Hydrofinishing to severe hydrotreating (for diesel)	Home heating oil, Diesel fuel, Turbine fuel	Nitrogen removal for stability, Cetane index improvement, Gravity improvement.
Fuel Oil	None to mild hydrotreating	Large stationary combustion fuel (e.g., steam boilers)	Upgrading (if any) dictated by end use.

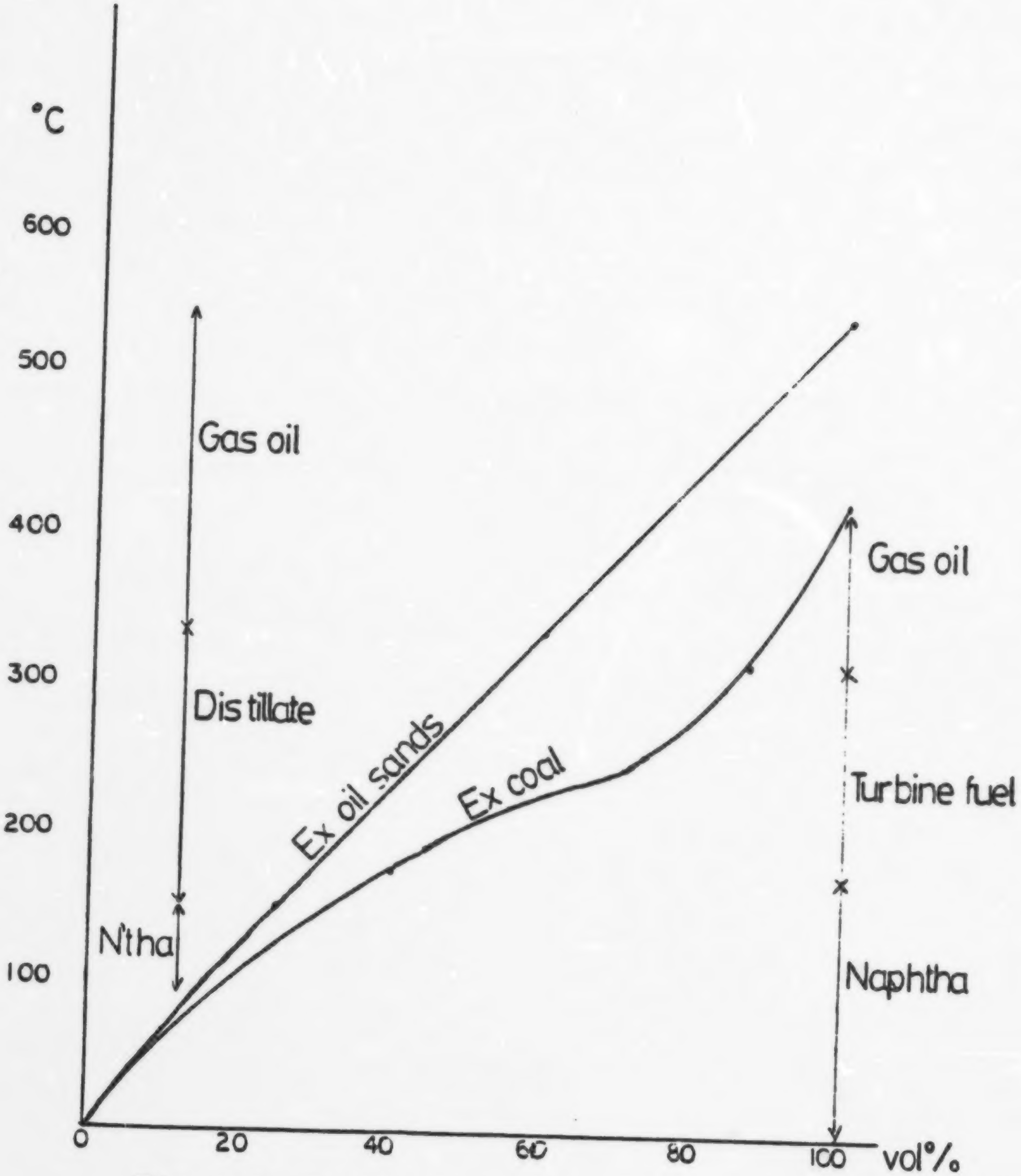


Figure6.1. Synthetic crude distillation analyses.

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8.0 APPENDIX - CALCULATIONS USED IN THE PREPARATION OF CAPITAL AND ANNUAL OPERATING COSTS

8.1 THE OIL SANDS PLANT

All costs for the oil sands plant were derived from the latest Alsands estimate (1). The calculations used for each cost detailed in section 5 of the main report are explained below.

8.1.1 Direct Field Costs

The following procedure was applied:

1) The split between the material and labour element of the cost was estimated (based on plnat complexity).
2) Each element was escalated from the mid 1978 Canadian dollars of the original estimate to mid 1982 Canadian dollars using factors of 1.5 for material costs and 1.4 for labour costs (see Table 1).
3) These costs were then corrected for capacity. The plant defined in this study has 87% of the Alsands capacity. A cost exponent of 0.6 was used to give an overall factor of:
   $0.87^{0.6} = 0.92$
   In the case of the surface mine only no correction was made for capacity because many expenses will be virtually unchanged by the small reduction in mine output. Those that would could easily be offset if for example oil sand of a lower average bitumen content were to be mined, necessitating a proportionately larger mine for a given synthetic crude output.
4) The separately escalated material and labour costs were recombined to give the direct field costs used in this report.

This calculation is summarized on Table 2. The calculation of the hydrogen plant cost requires further explanation. The Alsands source data gives a hydrogen synthesis plant cost of $181.4 million for an oil gasification unit with a subsequent $60 million to convert this to coke gasification. This study assumes a coke gasification unit at start up and a total cost of $241.4 million has been used in the calculation. This is clearly a conservative figure which does, however, allow a certain margin for cost escalation in that section of the plant. The calculated total cost of $326.4 million was split into $104.6 million for the oxygen plant and $221.8 for the hydrogen production units after calculating the oxygen plant cost using the same method that is described below for the H-Coal process.

8.1.2 Indirect Field Costs

Three such costs are given in the Alsands estimate:

Indirect costs	865
Home Office + fee	335
Contingency	200
Total	1 400

This total has been apportioned between the mine and the remaining operations in direct relation to their respective direct costs in the original source. Since most of these costs are labour related they have been escalated at a factor of 1.425 (closer to 1.4 then 1.5) to obtain 1982 values. The process cost has then been corrected for capacity as in the case of direct costs. Thus:

Mine	1400 x 0.27 x 1.425 x 1 =	538.7
Process	1400 x 0.73 x 1.425 x 0.92 =	1339.8
		1878.5

8.1.3 Preproduction Costs

The Alsands estimate identifies 22 items that contribute to the preproduction cost. These were individually estimated to require escalation of either 1.4 or 1.5 to convert from 1978 to 1982 dollars. These numbers were then corrected for capacity. Using this approach (summarized on Table 3) a pre-production cost of $918 million was calculated.

8.1.4 Post Start Up Costs

The Alsands estimate identifies two items. One of these, modifications to the gasification system, has already been incorporated into the direct costs in this study. The remaining item, conveyer belts, which will be required as the mine expands are escalated from 1978 to 1982 dollars with no capacity correction. Thus:

120 x 1.5 = 180

8.1.5 Working Capital

The Alsands estimate identifies three items. The first, spare parts, was calculated from the original estimate figure in exactly the same way as the indirect costs, i.e.:

Mine Spares = $29 \times 0.27 \times 1.5 \times 1 =$ 11.7
Process Spares = $29 \times 0.73 \times 1.5 \times 0.92 =$ $\underline{29.3}$
41.0

The two remaining items, chemical and oil inventories were considered to lie within the process area and calculated from the figures in the original estimate as follows:

Chemical inventory = $30.8 \times 1.5 \times 0.92 = 42.5$
Oil inventory = $13.5 \times 1.5 \times 0.92 = 18.6$

Total working capital is calculated.

Working capital = $41 + 42.5 + 19.6 =$ \$102 million

8.1.6 Interest During Construction

This was calculated by applying the formula:

$$IDC = ((1+i)^t - 1)\,PC$$

where

IDC = Interest during construction
i = Interest rate (as a fraction)
t = Time Period (years) from the mean point of project expenditure to project completion
PC = Plant Costs (Direct + Indirect)

In this case an interest rate of 15% (based on US Treasury Bills) was used. The project schedule was taken as 5 years with the mean point of expenditure as 3 years giving $t = 2$. Thus:

$$IDC = ((1.15)^2 - 1)(2767 + 1878) = \$1500 \text{ million}$$

8.1.7 Major Infrastructure Costs

The Alsands estimate includes provision for a new town, utilities corridor and airstrip at a total cost of \$642 million. These items are extremely site specific and no site has been specified for this study. However, to maintain a consistent approach the figures in the Alsands estimate were directly escalated to give a 1982 dollar estimate for use in this study. Thus:

642 x 1.5 = $963 million

8.1.8 Annual Operating Costs

The Alsands estimate identifies eight items. The first, wages and salaries, was given as $185 million/a rising by $10 million/a after coke gasification was introduced. The operating costs for this report were calculated by escalating these figures by 40% and correcting the portion associated with the process to reflect the smaller (87%) capacity. Thus:

Mines	185 x 0.27 x 1.40	69.9
Process	((185 x 0.73) + 10) x 1.40 x 0.92 =	186.8
		257.0 (round off)

Other operating costs are all material related and were escalated by 50% and, with the exception of one mining related item, corrected for capacity. The calculations are summarized on Table 4.

8.2 THE COAL MINE

Two principal sources of data were used. The major equipment items required were estimated from 2. Telephone quotes from several vendors were obtained for the capital costs of these items. These estimates were compared and average figures, summarized on Table 5, used in this study. An additional allowance was made for winterization. The remaining item costs and operating costs were estimated from a U.S. Bureau of Mines (USBM) report (3).

8.2.1 Direct Field Costs

Items other than those covered on Table 5 were directly estimated from the USBM report. The procedure used was:

1) Price escalation. The USBM report quotes 1979 U.S. dollars. To convert to 1982 these were escalated by 35%. A factor of 1.22 was used to convert from U.S. to Canadian dollars. The total escalation factor at constant capacity was thus:
   1.35 x 1.22 = 1.65
2) Capacity correction. The USBM report is based on a 5 million t/a mine. This study requires a 19 million t/a mine, a scale up of 3.8. Since the majority of the items required for the mine are large distinct pieces of equipment, a capacity correction

factor of 4 was applied with no exponent. The calculations are summarized on Table 6.

8.2.2 Indirect Field Costs

These were calculated from the direct field costs of $395 million using the following procedure which is defined in the USBM report.

Direct costs 395		
Field indirect 2% of direct field cost		7.9
Construction costs = 395 + 7.9		402.9
Engineering 2% of construction		8.1
Overhead and administration 5% of construction		20.1
	Subtotal	431.1
Contingency	15% of 431.1	64.7
Fee	10% of above indirects	
	10% (7.9 + 8.1 + 20.1 + 64.7)	10.1
	TOTAL INDIRECT COSTS	111

8.2.3 Preproduction Costs

For consistency, preproduction costs have been taken as the same percentage of mining direct costs as the mining items in the oil sands estimate (i.e. 14.8%). This figure was then corrected for the relative areas of the two mines (i.e. 1.5) and an additional 20% somewhat arbitrarily added to cover miscellaneous items. All other defined preproduction costs are included in the process section of the estimate. Thus:

395 x 0.148 x 1.5 x 1.2 = 105

8.2.4 Annual Operating Costs

Wages and Salaries - Wages and salaries were developed from manning charts presented in the USBM report for a 5 million t/a mine. To allow for the larger capacity of the mine in this study manning levels were increased generally in direct proportion (not by an exponent related factor since most mining operations are labour intensive). A few specific exceptions (e.g. the environmental and office managment staff) were increased at less than the full factor. Total hours worked per year were calculated for payroll staff and annual wages calculated at $25/h. Annual salaries were estimated for salaried staff. These calculations are summarized on Table 7 and yield an annual wage and salary cost of $34.8 million.

Other Operating Costs - These were calculated directly from figures presented in the USBM report. The figures calculated for maintenance and operating materials could well be a slight underestimate since Albertan winter conditions probably take a heavier toll of machinery than the American conditions on which the USBM report is based. As a proportion of total cost, however, the figure is so small that the overall estimate is insensitive to it. The calculations are summarized on Table 8. Thus the total operating cost is:

34.8 (Table 7)
42.9 (Table 8)
77.7

8.2.5 Coal Revenue

The coal output of the mine is 19.4 million t/a. Informal discussions with Manalta Coal (Calgary) indicated that the market price of sub-bituminous coal from a mine of this capacity should be $10 to $15 per short ton, i.e., $11 to $16.5 per tonne. For the sake of this study we have assumed a price of $15/t giving an annual mine operating revenue of 15 x 19.4 = $291 million/a.

8.2.6 Working Capital

This has been calculated adopting a procedure defined in the USBM report. The calculations are summarized on Table 9.

8.2.7 Interest During Construction

This was calculated by applying the formula:

$$IDC = ((1 + i)^t - 1) PC$$

where

IDC = Interest during construction
i = Interest rate (as a fraction).
t = Time Period (years) from the mean point of project expenditure to project completion
PC = Plant costs (Direct + Indirect)

In this case an interest rate of 15% (based on US Treasury Bills) was used. The project schedule was taken as 5 years with the mean point of expenditure as 3.5 years giving t = 1.5. Thus:

$$\mathrm{IDC} = (1.15^{1.5} - 1)(395 + 111) = 118$$

8.3 THE H-COAL LIQUEFACTION PROCESS

The report prepared by Fluor for EPRI (4) formed the basis of this estimate.

8.3.1 Direct Field Costs

These were calculated directly from the Fluor estimate. The production rate of liquid products from the plant in this report is 3.49 times that from the Fluor plant. However, because of changes made to the flowsheet either to accommodate the different coal feedstocks or to maintain a consistent approach with the oil sands estimate not all units were scaled up at the same rate. In prorating costs from the Fluor scale an exponent of 0.6 was used for conventional operations. For operations that are essentially multitrain (e.g. coal preparation) a 0.8 exponent was used. In the original Fluor estimate direct costs are clearly apportioned between the material and labour element. The material element was escalated from mid 1979 to mid 1982 dollars using a factor of 1.35, the labour element was escalated using a factor of 1.27 (see Table 1). To allow for contruction in Alberta a winterization element of 5% was added to the material element of the cost. The resultant figure was converted from U.S. to Canadian dollars by applying a 1.22 exchange rate.

The plant described in the Fluor report contained no hydrotreating facility so this had to be estimated independently. The estimate was based on the costs used for the oil sands pland (see Section 8.1 above). The two plants hydrotreat the same total throughput. Coal liquids are known to require more severe reaction conditions (higher pressure, and hydrogen demand). It seemed reasonable, therefore, to inflate the cost above that used in the oil sands estimate. In view of the lack of firm experiemental data on the particular coal liquids to be handled any cost estimate should be regarded as somewhat speculative. We applied an increase of 35% to the estimate used for the oil sands hydrotreaters. It should be noted that varying between a 0% and 100% increase only alters the total project direct cost estimate by 7%. Additional minor allowances were made in a number of areas. These and all the calculations described above are summarized on Table 10.

8.3.2 Indirect Costs

These were applied as follows:

Indirect field costs 44.2% of direct costs	1 298.1
Home office costs 5% of direct costs	146.8
Contingencies 15% of indirect field & home office costs	216.7
Fee 10% direct labour & indirect field & home office costs	213.2
	1 874.8

8.3.3 Preproduction Costs

These were matched to the same format used for the oil sands estimate. Some items were increased e.g. A larger allowance was made for plant changes because of the more involved technology. These items were more or less balanced by housing and relocation items that were decreased because of the plant's closer proximity to Edmonton. Preproduction costs for both the oil sands and liquefaction processes are detailed on Table 11.

8.3.4 Annual Operating Costs

Wages and Salaries - The EPRI report estimates that 82 operating jobs are reuqired. Assuming 4 shift coverage this gives 328 real jobs. At the study capacity (3.49 times EPRI capacity) this was estimated at:

$328 \times 3.49^{0.8} = 892$ jobs

The high exponent was used because of the large number of parallel trains in the coal preparation and liquefaction areas. Labour was priced at $25/h to give an annual process operating wage bill of:

$892 \times 25 \times 8760/4 =$ $48.8 million

(Each man assumed to be paid for 25% of the hours in a year).

An additional allowance of $4.3 million/a was made for the hydrotreaters which are not included in the EPRI report and a further allowance of 5% for recruiting/training. Thus a total:

$(48.8 + 4.3)\ 1.05 =$ $55.8 million/a

Other wages were calculated by using the same factor relative to operating labour defined in the EPRI report. Thus:

Maintenance labour = 55.8 x 1.73 =	96.5
Admin. labour & expenses = 55.8 x 2.46 =	137.3
	233.8

Thus the total wages bill is 233.8 + 55.8 = 289.6.

Materials - Maintenance materials were estimated as 50% above maintenance wages (as in the EPRI report). Thus:

1.5 x 96.5 = $145.1 million/a

Operating Materials - For consistency with the oil sands estimate these were taken as 31% of maintenance materials. Thus:

0.31 x 145.1 = 45.0

Chemicals and Catalysts - These were developed from a table presented in the EPRI report. Each catalyst or chemical cost was escalated from the source figure by correcting for relative capacity of that plant section, price escalation (35%) and currency exchange (22%). The calculations are summarized on Table 12. A further $5.3 million/a was added to allow for hydrotreating catalysts not covered in the EPRI table. The plant requires import of natural gas. (277950 t/a; 15 500 000 GJ/a) to the power plant. Canadian Western Natural Gas advised us that a price of $2.06/GJ should apply to give a natural gas charge of $32 million. Thus the total chemical charge is:

71.5 + 5.3 + 32 = $108.8 million/a

Corporate Overhead - Calculated as 10% of wages and salaries i.e., 10% of 289.6 = $29 million.

Property Tax and Insurance - For compatibility this was derived from the oil sands plant estimate as follows:

$$\text{PIT} = \text{Oil Sands Estimate} \times \frac{\text{Oil Sands Direct Costs}}{\text{Coal Liquefaction Direct Costs}} \times 15\%$$

The 15% is a somewhat arbitrary correction because the coal plant is closer to Edmonton. Thus:

$$\text{PIT} = 40 \times \frac{2936}{2767} \times 1.15 = 48.9$$

Imported Power - As in the oil sands case imported power is not required on a permanent basis but only to compensate for temporary imbalance. While the total power requirements of the coal plant are larger (2 - 3 times the oil sands plant), the large number of parallel trains should mean less pronounced swings in demand. We have therefore, assessed the power import at only 50% above that for the oil sands plant. Thus power cost:

3.9 x 1.5 = $5.9 million/a

Coal Import - Coal is charged to the liquefaction at the same price ($15/t) as it is credited to the mine. This gives an annual charge of $291 million.

8.3.5 Working Capital

For consistency with the oil sands estimate, the same three items were considered.

Oil inventory is the same and is priced as such. The other two inventory items are estimated by taking the same ratio of working capital to annual charge as was used in the oil sands estimate. Thus:

spare parts and warehouse inventory	40
chemicals/catalysts in inventory	107
oil inventory	18.6
TOTAL	165.6

8.3.6 Interest During Construction

This was calculated by applying the formula.

$IDC = ((1 + i)^t - 1)\,PC$

where

IDC = Interest during construction

i = Interest rate (as a fraction)

t = Time period (years) from the mean point of project expenditure to project completion

PC = Plant costs (direct + indirect)

In this case an interest rate of 15% (based on US Treasury Bills) was used. The project schedule was taken as 5 years with the mean point of expenditure as 3 years giving t = 2. Thus:

$IDC = ((1.15)^2 - 1)(2935.8 + 1874.8) = 1551.4$

8.4 THE COST OF UPGRADING COAL LIQUIDS

CANMET have expressed a particular interest in the portion of the total cost that should be attributed to the upgrading of the raw coal liquids. The technical and therefore, economic basis for estimating these costs is somewhat speculative as discussed in section 6 of the main report. It is a simple task to identify the elements of the total estimate that derive from the upgrading operation. It should be borne in mind that while any uncertainty in these costs has a small effect on the overall estimate, it has a more profound effect in this specific exercise.

8.4.1 Capital Costs

The direct field costs incurred by the upgrading plant are:

Hydrotreaters	333
Portion of hydrogen plant	33
Portion of utilities plant	20
	386

All other capital elements are calculated from the direct cost estimate and can therefore, be calculated from the ratio of upgrading to total direct costs as follows:

Indirect costs $\frac{386}{2936} \times 1875 = 246$

Preproduction costs $\frac{386}{2936} \times 782 = 102$

Working Capital $\frac{386}{2936} \times 166 = 22$

Interest during construction $(1.15^2 - 1)(386 + 246) = 209$

8.4.2 Operating Costs

Operating wages and salaries (specific amount added for upgrading 4.3 + 5%)	4.5
Chemicals and catalysts	5.3
Natural gas import (all saved as fuel gas consumed by steam reformer is released)	32
Property tax and insurance $\frac{386}{2936} \times 48.9$	6.4

All other operating costs are calculated from the operating wages and salaries as described in the coal liquefaction estimate. Thus:

Maintenance labour	1.72 x 4.5	7.8
Admin. labour and expenses	2.46 x 4.5	11.1
Maintenance materials	1.5 x 7.8	11.7
Operating materials	0.31 x 11.7	3.6
Corporate overhead	0.1 x (4.5 + 7.8 + 11.1)	2.3

Corporate Overhead 10% of 23.4	
PIT 386 x 48.9	2.3
2936	6.4
Imported power negligle.	
Coal import. Unchanged by upgrading.	
Working Capital	
386 x 165.6	
2936	22
IDC ($(1.15)^2$ - 1) (386 + 246)	209

Capital and Operating costs for upgrading are summarized in Table 13.

TABLE 1

ESCALATION FIGURES

	CAPITAL	LABOUR	REMARKS
1978	6%	6%	From mid 1978
1979	10%	8%	
1980	10%	8%	
1981	10%	8%	
1982	6%	5%	To mid 1982
OVERALL	50%	40%	

Table 2 Calculation of direct field costs - Oil Sands Plant

ITEM	SOURCE COST	MATERIAL/LABOUR SPLIT			MATERIAL/LABOUR COSTS $ 1982 AT ORIGINAL CAPACITY		MATERIAL/LABOUR COSTS $ 1982 AT STUDY CAPACITY		DIRECT FIELD COSTS USED IN STUDY
		%	$						
Mining	511.7	65/35	332.6	179.1	498.9	250.7	499.0	250.7	749.7
Extraction	154.6	65/35	100.5	54.1	150.7	75.7	138.7	69.7	208.4
Upgrading:									
:Diluent Recovery	35.7	70/30	25.0	10.7	37.5	15.0	34.5	13.8	48.3
:Primary Upgrading	190.0	70/30	133.0	57.0	199.5	79.8	183.5	73.4	256.9
:Hydrogen Synthesis	241.4	70/30	169.0	72.4	253.5	101.4	233.2	93.2	326.4
:Hydrotreating	182.9	70/30	128.0	54.9	192.0	76.9	176.6	70.7	247.3
:Gas Treating	23.3	70/30	16.3	7.0	24.5	9.8	22.5	9.0	31.5
:Sulphur Plant	29.0	70/30	20.3	8.7	30.5	12.2	28.0	11.2	39.2
Utilities	184.5	70/30	129.2	55.3	193.8	77.4	178.3	71.3	249.6
Offsites & Storage	452.0	65/35	293.8	158.2	440.7	221.5	405.4	203.7	609.1
TOTAL	1945						1899.7	866.7	2767

Source costs from (1)

All costs in millions of Canadian dollars.

Source cost of $452 million for offsites includes $12 million for a process computer.

TABLE 3

PREPRODUCTION COSTS

ITEM	ALSANDS ESTIMATE	ESCALATION FACTOR	1982 COST AT STUDY CAPACITY
Furniture	13.8	1.5	19.0
Roads	3.5	1.5	4.8
Technology	28.3	1.5	39.1
Catalysts	41.6	1.5	57.4
Environment	2.4	1.5	3.3
Insurance	12.5	1.5	17.2
82 Budget (79)	8.2	1.4	10.6
83 Budget (80)	21.7	1.4	27.9
Barge/Ferry	1.0	1.5	1.4
Changes	19.2	1.5	26.5
Operations Wages	109.9	1.4	141.5
Eng. Admin	34.0	1.4	43.8
Supplies	10.6	1.5	14.6
Attrition	12.3	1.5	17.0
Transportation	2.5	1.5	3.4
Housing	51.6	1.5	71.2
Training	6.0	1.4	7.7
Relocation	33.1	1.5	45.7
Start Up	16.7	1.4	21.5
Project Staff	103.6	1.4	133.4
Mining Programs	78.1	1.5	109.6
Start Up (Nat. Gas)	73.5	1.5	101.4
TOTAL COSTS	$ 676.3 mm		918.0 mm

TABLE 4

CALCULATION OF ANNUAL OPERATING COST

OIL SANDS PLANT

ITEM	ALSANDS ESTIMATE	1982 COST AT ORIGINAL CAPACITY	1982 COST AT STUDY CAPACITY
Wages & Salaries	195		257
Maintenance Materials	65	97.5	89.7
Operating Materials	20	30.0	27.6
Chemicals	21	31.5	29.0
Tax/Insurance	29	43.5	40.0
Reclaimation	4	6.0	6.0
Corporate Overhead	18	27.0	24.8
Miscellaneous	3	4.5	4.1
Total	355		478.0

TABLE 5. COAL MINE DIRECT COSTS. QUOTED ITEMS

ITEM	NO.	AVERAGE OF VENDOR QUOTES	WINTERISATION ALLOWANCE	TOTAL
Drag lines 53 m^2	4	114.0	5.0	119
Drag lines 44 m^2	5	130.0	5.5	135.5
Coal shovels 11 m^2	7	13.4	1.1	14.5
Front end loader	3	2.9	0.2	3.1
Trucks 100 t	36	30.6	2.3	32.9
TOTALS (quoted items)		290.9	14.1	305.0

TABLE 6. COAL MINE DIRECT COSTS. ESCALATED ITEMS

Item	Source Cost 1978 $US x 10^6	Capacity Factor	Winterisation Allowance	Cost at Study Capacity 1982 $ x 10^6 Cdn.
Dragline auxilliaries	0.351	9	0.2	5.4
Topsoil removal & reclamation	2.242	4	2.4	17.2
Maintenance equipment	0.856	4	0.2	5.8
Auxilliary equipment	8.694	4	2.9	60.3
				88.7
Quoted items (Table 5)				305.0
Total Direct Field Costs (round off)				395

NOTE: Capacity factor of 9 used for dragline auxilliaries based on number of draglines required.

TABLE 7. COAL MINE. WAGES AND SALARIES

DUTY	NO. OF PEOPLE	NO. OF DAY PER YEAR	HOURS/DAY	RATE	TOTAL
TOPSOIL & RECLAMATION					
Tractor Operators	48	220	8	25	2112000
Bulldozer Operators	48	220	8	25	2112000
	96				4224000
OVERBURDEN REMOVAL					
Bulldozer Operator	12	345	8	25	828000
Dragline Operator	27	345	8	25	1863000
Dragline Oiler	27	345	8	25	1863000
Dragline Groundsman	27	345	8	25	1863000
Dragline Mechanic	27	345	8	25	1863000
	120				8280000
COAL REMOVAL & LOADING					
Shovel Operator	14	220	8	25	616000
Shovel Oiler	14	220	8	25	616000
Front End Loader	6	220	8	25	264000
Driller	32	220	8	25	1408000
Driller Helper	32	220	8	25	1408000
Shooter	8	220	8	25	352000
Pitman	8	220	8	25	352000
Coal Haul Truck Drivers	54	220	8	25	2376000
	176				7744000
MAINTENANCE					
Road Grader Operator	16	220	8	25	704000
Water Truck Driver	16	220	8	25	704000
Fuel Truck Driver	8	220	8	25	352000
Lube Truck Driver	12	220	8	25	528000
Supply Truck Driver	8	220	8	25	352000
Mechanic Truck Driver	64	220	8	25	2816000

continues

TABLE 7 continued

DUTY	NO. OF PEOPLE	NO. OF DAY PER YEAR	HOURS/DAY	RATE	TOTAL
MAINTENANCE continued					
Electrician	40	220	8	25	1760000
Welder	40	220	8	25	1760000
Utility Man	60	220	8	25	2640000
	264				11616000
	656			TOTAL:	31.864.000

Salaried personnel are based on No. of craft people and are prorated from the Manning Study.

DUTY	NO. OF PEOPLE	YEARLY SALARY	TOTAL
Superintendant	4	64050	256200
Pit Foreman (gen)	4	43550	174200
Pit Foreman	12	35000	420000
Foreman (Maint)	12	35000	420000
Mining Engineer	4	49550	198200
Geologist	4	36700	146800
Environmentalist	2	33300	66600
Safety Inspectors	12	32450	389400
Office Managers	2	35000	70000
Purchasing Agents	4	35000	140000
Timekeepers	4	24000	96000
Bookeepers	4	24000	96000
Warehousemen	12	22200	266400
			2931000

Salary is based on costs from Manning x 1.40 Labour Factor x 1.22 Exchange. (Rounded off)

TOTAL: $34.795.000.

TABLE 8. COAL MINE. ANNUAL OPERATING COSTS (OTHER THAN WAGES)

ITEM	SOURCE COST	STUDY COST
Power	1.6	10.0
Operating Materials	2.13	7.8
Maintenance Materials	2.12	7.8
Corporate Overhead	1.23	4.5
Property Taxes & Insurance	1.98	7.3
Reclamation & Revegetation	1.5	5.5
TOTAL		42.9

All calculations (except power).

$$\text{Study cost} = \text{Source cost} \times \underset{\text{escalation}}{1.35} \times \underset{\text{US exchange}}{1.22} \times \underset{\text{Capacity}}{3.8^{0.6}}$$

$$= \text{Source cost} \times 3.67$$

Power is corrected linearly not exponentially for capacity.

TABLE 9. COAL MINE. WORKING CAPITAL

ITEM	CALCULATION	TOTAL
Direct Labour - 4 man ths wages & salaries	0.33 x 34.8	12.0
Operating Supplies	0.33 x 15.6	5.2
Payroll Overhead 38% of Direct Labour	0.38 x 12	4.6
Indirect Costs 1% of D.L.	0.01 x 395	4.0
Fixed Costs 0.5% of D.L.	0.005 x 395	2.0
Spare Parts 1% of D.L.	0.01 x 395	4.0
Miscellaneous 0.1% of D.L.	0.001 x 395	0.4
	TOTAL	32.2

TABLE 10. CALCULATION OF DIRECT FIELD COSTS. H-COAL PROCESS

ITEM	SOURCE COST $US 1978		MATERIAL/LABOUR COSTS $CDN. 1982 @ ORIGINAL CAPACITY		RELATIVE CAPACITY STUDY - SOURCE	COST EXPONENT	MATERIAL/LABOUR COSTS $CDN. 1982 @ STUDY CAPACITY		ADDITIONAL ITEMS		REMARKS	COSTS USED-STUDY		TOTAL COSTS USED IN STUDY
	Material	Labour							Material	Labour		Material	Labour	
Coal Preparation	29.072	11.779	50.276	18.250	3.49	0.8	135.8	49.8	7.0	-	Note 1	142.8	49.8	192.6
Liquefaction	57.090	49.431	271.664	76.588	3.49	0.8	738.2	208.4	42.0	-	Note 1	780.2	208.4	988.6
Oxygen Plant	27.741	6.515	47.974	10.094	3.49	0.6	101.2	21.3	55.5	33.1	Note 2	156.7	54.4	211.1
Hydrogen Plant:														528.4
Gasification	89.253	23.909	154.350	37.045	3.49	0.6	326.1	78.2	-	-	-	326.1	78.2	
Reforming	20.646	4.769	35.704	7.389	5.82	0.6	102.8	21.3	-	-	-	102.8	21.3	
Hydrotreating									241.1	91.9	Note 3	241.1	91.9	333.0
Gas Treating	5.111	1.646	8.839	2.550	3.49	0.6	18.6	5.5	-	-	-	18.6	5.5	24.1
Sulphur Plant	4.753	1.383	8.22	2.143	1.83	0.6	11.8	3.1	3.2	3.4	Note 2	15.0	6.5	21.5
Utilities	63.34	20.715	109.537	32.096	3.49	0.6	231.4	68.9	20	-	Note 4	251.4	68.9	320.3
Offsites	43.480	19.633	75.192	30.419	3.49	0.6	158.65	64.2	16	14	Notes2&5	174.6	78.2	252.8
Storage	6.091	4.245	10.533	6.577	5.18	0.8	39.3	24.1	-	-	-	39.3	24.1	63.4
TOTAL												2248.6	587.2	2935.8

NOTES

1) Nominal 5% allowance to material costs for additional intermediate storage to handle the large number of parallel trains in these areas.

2) These cover additional subcontract items defined in the EPRI report.

3) Taken as 35% above oil sands case.

4) Nominal $20million allowance for additonal load on part of the utilities system imposed by the hydrotreaters.

5) Offsites here includes effluent treatment costs defined in the EPRI report.

TABLE 11. PREPRODUCTION COSTS

ITEM	OIL SANDS		COAL LIQUEFACTION
Furniture	19.0		19.0
Roads	4.8		5.0
Technology	39.1		30.1
Catalysts	57.4		51.2
Environment	3.3		3.5
Insurance	17.2		20.8
82 Budget (79)	10.6		13.9
83 Budget (80)	27.9		34.7
Barge/Ferry	1.4		N.A.
Changes	26.5		69.5
Operation Wages	141.5		77.2
Eng. Admin	43.8		55.6
Supplies	14.6		17.6
Attrition	17.0		20.8
Transportation	3.4		3.5
Housing	71.2		21.4
Training	7.7		7.7
Relocation	45.7		36.6
Start Up	21.5		25.8
Project Staff	133.4		146.7
Mining Programs	109.6		N.A.
Start Up (Nat. Gas)	101.4		121.4
TOTAL COSTS	918.0	mm	782.1

TABLE 12. COAL LIQUEFACTION. CHEMICAL AND CATALYST CHARGES (EXCLUDING HYDROTREATERS)

UNIT	CHEMICALS/CATALYSTS	COST AT EPRI CAPACITY 1978$US x 10^6	CAPACITY RATIO (STUDY CAPACITY) EPRI CAPACITY	COST AT STUDY CAPACITY 1982$CDN x$10^6$
200	American Cyanamid	10.49	3.49	60.3
300	DGA(Diglycolamine)	0.017	3.49	0.1
400	Sour shift catalyst	0.624	3.49	3.6
	Selexol solvent	0.126	3.49	0.7
	HDS catalyst	0.050	5.82	0.5
	Zinc Oxide Adsorbent	0.156	5.82	1.5
	Reforming Catalyst	0.080	5.82	0.8
	Hi-Temp Shift Catalyst	0.024	5.82	0.2
	Low-Temp Shift Catalyst	0.100	5.82	1.0
	Potassium Carbonate	0.005	5.82	0.1
	Gas Treating Chemicals	0.004	5.82	0.1
600	Activated Alumina	0.005	1.83	0.02
	Cobalt Moly Catalyst	0.006	1.83	0.02
	Stretford Process Chemicals	0.020	1.83	0.06
700	Solvent (for tar acids)	0.021	3.49	0.1
	Sour Water Stripper Chemicals	0.018	3.49	0.1
900	Lime	0.064	3.49	0.4
	Soda Ash	0.098	3.49	0.6
	Sulphuric Acid	0.091	3.49	0.5
	Caustic Soda	0.020	3.49	0.1
	Cooling Water Chemicals	0.102	3.49	0.6
	Boiler Feedwater Chemicals	0.014	3.49	0.1
TOTAL				71.5

TABLE 13. COAL LIQUIDS UPGRADING CAPITAL AND ANNUAL OPERATING COSTS ($ x 10^6)

CAPITAL

Direct Field Costs	386
Indirect Field Costs	246
Preproduction Costs	103
Working Capital	22
Interest during Construction	209
TOTAL	966

OPERATING (per annum)

Wages and Salaries	23.4
Operating materials, chemicals	40.9
Maintenance materials	11.7
Corporate Overhead	2.3
Property Tax and Insurance	6.4
TOTAL	84.7

REFERENCES

1. Addendum to ERCB report 79- 1. (This contains supplemental information by Alsands including a 1979 update of capital and operating costs). January, 1980.

2. "Proposed keephills thermal station expansion. Units 3 and 4." Calgary Power Ltd. ERCB application no. 790693. October 1979.

3. "Basic estimated capital investment and operating costs for sub-bituminous and lignite coal strip mines". L.H. Berkshire. NTIS report no. FE/EES-79/6. April 1979.

4. "Engineering evaluation of conceptual coal conversion plant using the H-Coal liquefaction process". Fluor Engineers and Constructors Ltd. EPRI report No.AF-1297. December 1979.